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COST AND PERFORMANCE EVALUATION OF IN-PLANT TRIHALOMETHANE CONTROL TECHNIQUES

by

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Cincinnati, Ohio 45268

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16. ABSTRACT

Bench-, pilot-, or full-scale studies were conducted to evaluate the cost and performance of selected treatment technologies for reducing trihalomethane concentrations in drinking water. The four Florida plant sites that were selected for study used highly organic surface or ground raw water sources and served populations of less than 30,000. Low-pressure membrane processes (ultrafiltration), polyvalent aluminum chloride (PACl) coagulation, dissolved air and froth flotation, lime softening followed by alum coagulation, and modifications to conventional lime softening and alum coagulation were investigated. Short-term pilot-scale testing of an ultrafiltration process showed great promise as a THM control technology; however, further work on process pretreatment and disposal of reject water is indicated.

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FOREWORD

The U.S. Environmental Protection Agency is charged by Congress with protecting the Nation's land, air, and water systems. Under a mandate of national environmental laws, the agency strives to formulate and implement actions leading to a compatible balance between human activities and the ability of natural systems to support and nurture life. The Clean Water Act, the Safe Drinking Water Act, and the Toxic Substances Control Act are three of the major congressional laws that provide the framework for restoring and maintaining the integrity of our Nation's water, for preserving and enhancing the water we drink, and for protecting the environment from toxic substances. These laws direct the EPA to perform research to define our environmental problems, measure the impacts, and search for solutions.

The Water Engineering Research Laboratory is that component of EPA's Research and Development program concerned with preventing, treating, and managing municipal and industrial wastewater discharges; establishing practices to control and remove contaminants from drinking water and to prevent its deterioration during storage and distribution; and assessing the nature and controllability of releases of toxic substances to the air, water, and land from manufacturing processes and subsequent product uses. This publication is one of the products of that research and provides a vital communication link between the researcher and the user community.

The control of chronically toxic organics in drinking water is currently a major concern in drinking water treatment. Conventional and newly developed water treatment processes are being evaluated for reduction or removal of organic contaminants. This project investigated the membrane processes, polyvalent aluminum chloride (PACl) and alum coagulation, lime softening, and flotation on a bench-, pilot- and/or plant-scale for their capability to meet the THM MCL. Actual cost and performance were documented on a plant scale for PACl and alum plus lime softening.

Francis T. Mayo, Director
Water Engineering Research Laboratory

ABSTRACT

This project investigates the performance and costs of controlling trihalomethane (THM's) in water treatment plants with techniques emphasizing new technology at project sites serving fewer than 75,000 consumers. The sites for this project were selected after consultation with EPA Region IV. They included the Olga water treatment plant in Lee County, Florida, the Venice water treatment plant in Venice, Florida, the ACME Improvement District (AID) water treatment plant in West Palm Beach, Florida, and the Village of Golf (VOG) water treatment plant in the Village of Golf, Florida. Technologies investigated for THM reduction were optimization of lime-softening and alum coagulation plants, polyvalent aluminum chloride (PACl) coagulation, lime-softening and alum coagulation in series, dissolved air flotation and froth flotation, and membrane processes.

Plant optimization of lime softening plants involved increasing the softening reaction pH to 11.0 to 11.3, adding H_2SO_4 to offset the high reaction pH before filtration, and reducing the chlorine dose because of the decreased organic demand. This technique reduced THMs by an average of 30% to 40%, increased final hardness by 80 mg,/L $CaCO_3$, and increased O&M costs by $0.02/1000 gal or 3% to $0.22/1000 gal or 17%. Venice THM's were reduced to the 0.1-mg/L maximum contaminant level (MCL) by this technique. Optimization of an alum coagulation plant during the wet and dry seasons involved controlling the coagulation pH at 5.5 to 6.5 with alum or H_2SO_4 and dosing with adequate alum (80 to 120 mg/L) to remove 60% to 70% of the DOC. Reaction pH was more significant than coagulation dose for the plant test conditions. The least chemical cost involved using alum only to remove organics and H_2SO_4 to decrease reaction pH. THM formation was about 15% higher during the wet season. Plant O&M costs were increased $0.09/1000 gal or 8% during the wet season and $0.18/1000 gal or 30% during the dry season. The average THM reduction during the wet and dry seasons were 72% and 23%, respectively but the MCL was always exceeded. Two PACl coagulants were investigated--100S, a sulfate chloride base, and 190, a chloride base, PACL 100S was superior for organic reduction. Plant tests with PACl 100S showed that it, like alum, was dependent on reaction pH and dose for organic removal. PACl is far less acidic than alum and required the use of acid to achieve optimum reaction pH. Removals of color, DOC, THMPF, and TOXFP were generally slightly less than those for alum (5%). Plant tests of PACl 100S increased O&M costs by $0.55/1000 gal or 51% during the wet season and $0.42/1000 gal or 70% during the dry season. The corresponding THM reduction was 69% during the wet season and 29% during the dry season but the THM MCL was exceeded.

One reverse osmosis and six ultrafiltration membranes were investigated for THM precursor reduction. These investigations showed that an ultrafilter rejecting molecular weights of 2000 and higher will remove 50% of the DOC and still exceed the THM MCL. However, an ultrafilter capable of rejecting molecular weights of 400 and higher was found to remove 90+% of the DOC and met the THM MCL at three different sites which originally had THM's in excess of twice the MCL. Subsequent to bench-scale testing, a 25,000-gpd ultrafiltration membrane pilot plant was built and operated at two different sites for approximately 5 months. Feed pressures of 100 psi and recoveries of 95% continually met the THM MCL and reduced hardness by 40%. Estimated cost was $1.00/gal for construction and $0.47/1000 gal for O&M costs for a 1 mgd membrane plant.

Pilot and plant studies at 0.25 MGD indicated that sequentially treating water with lime-softening and alum coagulation would not remove any more THM precursors than would alum coagulation alone. Dissolved flotation following alum coagulation offerred no additional THM precursor removal relative to alum alone. The DOC increased with foam flotation of raw water using alcohol surfactants common to the mining industry.

This report was submitted in fulfillment of Cooperative Agreement No. CR-811039-01-0 by the University of Central Florida under the sponsorship of the U.S. Environmental Protection Agency. This report covers the period 9/13/83 to 9/13/85, and work was completed as of 9/13/85.

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ABBREVIATIONS AND SYMBOLS

1. AID - Acme Improvement District
2. Alk - total alkalinity
3. Alum - $Al_2(SO_4)_3 \cdot 14H_2O$
4. BLS - Bureau of Labor Statistics
5. BTI - Basic Technologies, Inc.
6. BW30 - Filmtec BW 3030 Brackish Water Membrane
7. °C - Celsius
8. CaH - calcium hardness
9. cp - centipoise
10. CI - confidence interval
11. cpu - chloroplatinate units
12. ΔP - change in pressure
13. DeSal - Desalination, Inc.
14. DPD - n-n-diethyl-p-phenylenediamine
15. DOC - dissolved organic carbon
16. EDTA - ethylene diaminetetra acetic acid
17. ENR - Engineering News Record
18. EPA - U.S. Environmental Protection Agnecy
19. °F - Fahrenheit
20. FDER - Florida Department of Environmental Regulation
21. FL - Florida
22. ft^2 - feet squared
23. FW - flux of water
24. GAC - granular activated carbon
25. GF/A - glass fiber filters
26. gm - grams
27. gpd - gallons per day
28. gpm - gallons per minute
29. > - greater than
30. ≥ - greater than or equal to
31. hr - hour
32. HRS - Health and Rehabilitative Services
33. I - industrial statistic
34. Inst THM - instantaneous trihalomethane
35. L - liter
36. lb - pound
37. < - less than
38. ≤ - less than or equal to
39. L.S. - lime softening
40. MCL - maximum contaminant level
41. meq - milliequivalent

42. mg - milligrams
43. MGD - million gallons per day
44. MgH - magnesium hardness
45. ml - milliliter
46. mls/min - milliliters per minute
47. MW - molecular weight
48. NCH - noncarbonate hardness
49. N50 - Filmtec Nanofilter membrane
50. NM - nanometer
51. NTU - nephelometric turbidity units
52. Osmo PT-2 - Osmontics Osmo PT-2 Nanofiltration Membrane
53. Osmo PT-4 - Osmontics Osmo PT-4 Nanofiltration Membrane
54. P - transmembranic pressure
55. PACl - polyvalent aluminum chloride
56. pH - $-\log[H_3O^+]$
57. π - difference in osmotic pressure
58. ppb - parts per billion
59. ppm - parts per million
60. psi - pounds of force per square inch
61. psig - gage pressure in pounds per square inch
62. PVC - polyvinyl chloride
63. R.O. - reverse osmosis
64. rpm - revolutions per minute
65. s - sample standard deviation
66. sec - seconds
67. SLR - surface loading rate
68. STP - sodium tripolyphosphate
69. TermTHM - terminal trihalomethanes
70. TDS - total dissolved solids
71. TH - total hardness
72. THM - trihalomethane
73. THMFP - trihalomethane formation potential
74. TOC - total organic carbon
75. TOX - total organic halides
76. TOXFP - total organic halides formation potential
77. TSP - trisodium polyphosphate
78. TSS - total suspended solids
79. UCF - University of Central Florida
80. ucl - upper control limit
81. μ - micron
82. U.F. - ultra filtration
83. μg - microgram
84. UFP 4040 - Desal UFP 4040 Ultrafiltration Membrane
85. U90-G10 - Desal U90-GIO Ultrafiltration Membrane
86. U90-G50 - Desal U90-G50 Ultrafiltration Membrane
87. uwl - upper warning limit
88. VOG - Village of Golf
89. Wp - coefficient of water permeation
90. WTP - water treatment plant
91. $\bar{x}$ - mean

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SECTION 1

INTRODUCTION

1.0 General

Many process developments occurred in the late 1970's and early 1980's to control trihalomethanes (THM's) in drinking water to the 0.10-mg/L maximum containment level (MCL). THM's were shown to exist in virtually every public water supply that used chlorine in the treatment process by the National Organics Reconnaissance Survey (Symons 1978). Many suppliers have had to alter their treatment process to conform to the THM MCL. Information concering the cost of implementing such processes is highly desirable and is required information in most cases before such changes are undertaken.

Some documentation is available that summarizes THM control. "Treatment Techniques For Controlling Trihalomethanes in Drinking Water" has been published both by the U. S. Environmental Protection Agency (EPA 1981) and the American Water Works Association (AWWA 1982). Gordon Culp has edited "Trihalomethane Reduction in Drinking Water," which is a text recently published by Noyes Data Corporation (Culp 1984). These publications summarize the capability of various unit operations to remove THM precursors and THM's and/or to form THM's. Existing data on the effect of unit operations on precursor or THM removal is incomplete, and a continuing need exists to provide such information to the field. This report is intended to provide new information on the cost and performance evaluation of in-plant THM control techniques.

1.1 Scope of Work

The scope of work of this cooperative agreement was to (1) identify new technologies for THM control, (2) identify sites where in-plant studies for cost and performance of THM control were needed and possible, and (3) conduct pilot plant and/or in-plant studies for cost and performance of THM control. The sites selected for in-plant work were limited to those serving a population of 75,000 or fewer residents, with a special emphasis on those potable water systems serving fewer than 10,000.

1.2 Selection of New Technologies

The available techniques for THM control are summarized in Table 1. These techniques are categorized as (1) THM removal, (2) THM precursor removal, (3) alternative disinfectants, and (4) changing water source. As

Table 1. Techniques to Meet the Trihalomethane Regulation

1. THM Removal	1.1	Oxidation: Other than Chlorine, primarily O_3, ClO_2, $KMNO_4$ or possibly ultraviolet radiation and oxidant combinations.
	1.2	Aeration: Such as diffused air, packed-tower, or spray aeration.
	1.3	Adsorption: Generally powdered activated carbon or granular activated carbon adsorption, including biologically activated carbon.
	1.4	Resins: Ion exchange resins.
2. THM Precursor Removal	2.1	Oxidation: Same as 1.1 only oxidation point is before chlorination.
	2.2	Adsorption: Same as 1.3, only point of carbon application is before chlorination.
	2.3	Coagulation: Adding or optimizing coagulation for precursor removal, includes filtration.
	2.4	Resin: Adding resin for precursor removal during coagulation, softening or in a separate clarification operation for precursor removal.
3. Alternative Disinfectants	3.1	Use of alternative disinfectant to reduce or eliminate the chlorine dose in existing treatment train. Includes moving point of chlorination in existing treatment train.
4. Changing Raw Water Source	4.1	Changing from a raw water source that contains a large amount of precursor to one that contains a small amount of precursor. Typically changing from a shallow aquifer or a surface water source to a deep well source.

mentioned earlier, much work had been accomplished for cost and performance documentation of THM control, and technologies not considered for this project were:

1. Movement of chlorination point
2. Chloramines
3. Chlorine dioxide
4. Granular activated carbon.

Other technologies that were considered for this project were:

1. Improved clarification
2. Powdered activated carbon
3. Membrane processes
4. Aeration
5. Ozonation with and without ultraviolet light
6. Installation of conventional treatment where none exists
7. Polymeric resins
8. Flotation.

The THM control technologies are not meant to be inclusive of Table 1. From the available technologies, coagulation with polyvalent aluminum chloride (PACl), membrane processes, sequential precipitation, and flotation were selected as the new technologies to investigate for cost and performance of THM control. These processes were selected because of adaptability to existing plants, availability of pilot plant process equipment, and lack of information on how THM control is effected by these processes. Ozonation in combination with ultraviolet radiation was a possible technique, but projects using these processes were not completed, which would have reduced the effectiveness of an in-plant project at this time. Other technologies may also have been appropriate for in-plant studies, but PACl coagulation, membrane processes, flotation, and sequential treatment were judged to present the most effective technologies to investigate.

1.3 Selection of Sites

The selection of sites for the study was national in scope, but because of budget considerations, it was limited to the southeast United States in EPA Region IV. An official inquiry was made to Region IV headquarters in Atlanta requesting any information on plants serving fewer than 75,000 residents. Emphasis was placed on plants that (1) served fewer than 10,000 residents, (2) had THM's in excess of the 0.10-mg/l MCL, (3) could use technical aid, and (4) wanted a solution to the THM problem. (Facilities serving more than 75,000 residents had already instigated solutions for THM control at this time, if necessary.) No formal information existed at that time but after a long discussion with Mr. D. Guinyard, Chief of the Water Supply Branch of Region IV, it was decided that the abundance of THM problems in Florida and the budget, made a Florida site superior to any other in Region IV.

The Florida Department of Environmental Regulation (FDER) was contacted and requested to provide any information available on plants with a service population of fewer than 75,000. FDER responded by providing the data shown in Table 2, which lists water plants in Florida serving more than 40,000. The notation designating THM problems was made after contacting each of the 12 FDER district offices and requesting THM information. In Florida, 5,453 drinking water treatment plants serve a total population of 12 million. There are 26 plants (0.5 percent) that each serve populations in excess of 75,000. These large plants serve a total population of 6 million. There are 146 plants (3 percent) that serve populations of 10,000 to 75,000, for a total service population of 3.6 million. The remaining population (2.3 million) is served in small groups of less than 10,000 by 5,281 (96.5 percent) treatment plants. Approximately 100 membrane (RO) plants are located in Florida, mostly serving populations of less than 10,000. Although few of these plants measure THM's, none have reported THM's in excess of the 0.10-mg/L MCL. Finally, the Florida Department of Health and Rehabilitative Services (FDHRS) was also contacted and requested to provide any available information on THM problems.

The Florida sites were then ranked by 10 factors as shown in Table 3. Although each of the sites was exceptional in its own right, the chosen sites were the Lee County Olga plant, Venice, ACME Improvement District (AID), and the Village of Golf (VOG).

1.4 Technology and Site Correlation

The four technologies to investigate were sequential precipitation, membrane processes, PACl coagulation, and flotation. Olga and AID both had process equipment that could support sequential precipitation (alum coagulation and lime softening), so those sites were chosen for these processes. Olga commonly used Alum to treat a highly colored surface water, so PACl coagulation was also assigned to Olga. VOG and AID are both small facilities that softened moderately hard water and have limited personnel, so membrane processes were assigned to those sites. Venice had both an RO plant and a lime-softening plant that produced THM's only slightly above the 0.1-mg/L MCL. Thus monitoring of their processes for THM's and manipulation of the combination of their processes was the only investigation conducted at Venice. No plant had any flotation facilities. A pilot flotation plant was available from Komline Sanderson for dissolved air flotation (DAF), which could be handled at the facilities at the Olga plant and stored in a trailer. The highly organic water at Olga was believed to be more suitable to flotation, so it was decided to investigate both DAF and froth or foam flotation (FF) at Olga. Because of the success of the membrane work at AID and VOG, a small-scale membrane unit was operated at Olga. These site/technology correlations are summarized in Table 4. Finally, it was necessary to optimize each process for THM precursor removal to have an equitable base for comparison of new technology.

Table 2. Water Plants in Florida Serving Populations Greater Than 40,000 and Designated for THM Problems

ID#	Location	Population
4131202	Rex Utilities Inc-Leisure City	000040000
6531992	Winter Haven Water Department	000040000
4060167	Broward Co - 1A	000041500
5360170	*Lee County Utilities Department of Water	000044000
5364048	*Lee County Utilities - South	000045000
2160816	Norwood	000046520
4060163	Broward Co - 2A	000048000
4130873	MDWASA-Carol City	000048000
2160322	Fairfax	000048300
6521047	Largo Water Department	000050000
6580326	Sarasota-City of	000050000
4061083	*Pembroke Pines	000050500
5110198	Naples Water Department	000051622
4500145	**Boynton Beach, City of	000055000
5440093	FKAA Lime Softening Plant	000057500
4060487	Fort Lauderdale, Dixie	000059000
4501124	*Seacoast Utilities	000059363
4130969	**No. Miami Beach County of Dade	000059653
3481482	Winter Park, City of	000063000
4060845	Margate, City of	000065000
4500130	**Boca Raton, City of	000065000
4130977	**North Miami, City of	000065903
5444047	FKAA Key West R. O.	000066000
4130968	**No Miami Beach, Sunny Isles	000067800
3640275	*Daytona Beach, City of	000069073
2160488	Highlands	000069200
2160640	Lake Shore	000070410
4061129	Pompano Beach, City of	000070500
2010946	Gainesville-Ala. County Reg. Utilities	000075000
6520336	Clearwater Public Water Utilities	000075000
2160736	McDuff S/D	000079660
4501559	**West Palm Beach, City of	000084000
6531014	Lakeland Department of Electricity & Water	000090000
2160710	Main Street	000095210
3050223	Cocoa, City of	000103849
3051447	*Melbourne, City of	000104062

*Definite 1982-83 THM Problem

**Probable 1982-83 THM Problem

continued

Table 2 (continued). Water Plants in Florida Serving Populations Greater Than 40,000 and Designated for THM Problems

ID#	Location	Population
6411132	Manatee County Utility Department	000105000
4060642	Hollywood, City of	000112000
1370655	Tallahassee, City of	000121000
4130901	*Miami Beach, City of	000125000
4431792	Southern Gulf Utilities	000135000
2161328	Jacksonville-South Grid	000136220
4130604	Hialeah, City of	000142000
4130524	General Waterworks South Florida District	000142200
4131613	General Waterworks of South Florida-Consumers	000174580
4060486	Fort Lauderdale, City of	000181912
1170525	Escambia County Utility Authority	000196000
4131618	North Miami Beach	000200000
41308755	*MDWASA-Hialeah Preston System	000243147
3480962	Orlando, City of	000283480
6521715	St. Petersburg Water Department	000290000
6512230	West Regional Water Supply Authority	000300000
2161327	Jacksonville-North Grid	000394750
6290327	*Tampa City Water Department	000475000
4131617	MDWASA-Palmetto Country Club	000560000
4130871A	MDWASA-Main System	00107500

*Definite 1982-83 THM Problem

**Probable 1982-83 THM Problem

Table 3. Site Rating Table

Item	Village of Golf	Port St. Joe	Pohokee	Holly Hill	ACME	Venice	Jupiter	Lee County	Bay City	Ft. Myers
Service Population	2,500	5,200	6,000	10,000	10,000	14,000	20,000	18,000	30,000	35,000
Supply	Well	Well	Surface	Well	Shallow Well	Well	Well	Surface	Surface	Well
Variable Quality	No	No	Highly	No	Moderately	No	No	Highly	Highly	No
Term THM	450	—	800	300	400	130	50	900	130	300
Cost Records	Good	Fair	Fair	Good	Good	Good	Good	Good	Good	Good
Cooperation	Good	Fair	Good	Excellent	Excellent	Very Good	Fair	Fair	Excellent	Good
Unused Equipment	No	No	No	No	Yes	No	Yes	No	No	No
Location	Good	Fair	Good	Excellent	Good	Good	Good	Excellent	Good	Good
Compliance Required	No	No	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Laboratory	Basic	Basic	Basic	Certified	Certified	Certified	Certified	Excellent	Certified	Certified
Other				Rapid THM Formation	Money Complete 0.6-mgd Pilot Plant Available	Money Membrane and Lime Softening Plant	No THM Problem	2 Plants- (A) ClO_2, GAC Coag. and Softening; (B)Lime Softening Both with THM Problem		High Conc. Br Species

Table 4. Site/Technology Correlation

Site	Technology	Scale
Olga	Plant Optimization	plant
	PACl Coagulation	plant
	Flotation	bench
	Membranes	pilot
	Precipitative Processes	plant
VOG	Plant Optimization	plant
	Membranes	pilot
AID	Plant Optimization	plant
	Membranes	pilot
	Precipitative Processes	plant
Venice	Plant Optimization	plant

SECTION 2

CONCLUSIONS

2.0 Plant Optimization

1) DOC, THMFP, TOXFP, and color are reduced as the reaction pH during softening is increased.

2) If the THM's are 150 µg/L or less, it is probable the THM and Cl_2 residual MCL can be met in a lime-softening plant by increasing the reaction pH and lowering the Cl_2 dose.

3) Optimizing a lime-softening plant for THM control will increase product hardness by 75 mg/L as $CaCO_3$ or more.

2.1 PACl/ALUM Coagulation

1) PACl has less base neutralizing capacity than alum, which would generally be a disadvantage for organic removal from a low-turbidity, high-color water because of the importance of a low-coagulation pH.

2) Color, DOC, THMFP and TOXFP reduction by either PACl or alum coagulation was maximized near pH 5.5.

3) Aluminum residuals after PACl or alum coagulation dependend on coagulation pH and were minimized from pH 5.0 to 6.0.

4) DOC was a better surrogate parameter for THM and TOX than color.

5) Unit cost was more dependent on seasonal water demand than chemicals required for optimum organic removal.

6) Preceding either PACl or alum coagulation with lime softening does not increase the organic removal during coagulation.

7) THM:TOX ratios were less variable than DOC:THM:TOX because of the effect of chlorine dose on THM and TOX.

8) Alum floc was more settleable at low coagulant doses (80 mg/L) than PACl floc.

9) Alum was slightly superior to PACl for color, DOC, and THMFP reduction.

2.2 Sequential Treatment

1) The order of lime softening and alum coagulation was not significant for maximizing organic removal.

2) Alum coagulation when combined with a lime softening process will significantly increase the organic removal and reduce the THMFP. However, lime softening will not increase the organic removal when combined with an alum coagulation process.

2.3 Membranes

1) A 150-day pilot plant UF of two highly organic ground waters (which produced THM's above 400 µg/L when conventionally treated) produced a finished water that easily met the Cl_2 residual and THM MCL's.

2) Compared with membranes, no other tested process has the same practical capacity for the removal of THM precursors.

3) The vast majority of THM precursors were rejected by a membrane that has a molecular weight cutoff of less than 2000 but greater than 400.

4) The effect of percentage of recovery and feed pressure on product water quality was site-specific.

5) Flux was site-specific ranging from approximately 20 to 10 gpd/ft^2, and it was generally independent of the percentage of recovery.

6) Caloosahatchee River water needs more elaborate pretreatment than single-pass sand filtration to maintain N-50 permeate flux. Flux values during the extended study were 12 to 18 gpd/ft^2, and they decreased continuously with time of operation.

7) N-50 permeate quality is a function of feed pressure, percentage of recovery, and membrane surface condition. The best inorganic and organic rejection occurred at the highest pressures, the lowest recoveries, and the cleanest membrane condition. With the Caloosahatchee water, the THM MCL was met only at the 105 psi, 60% recovery and the 75 psi, 60% recovery conditions for a cleaned membrane. Membrane surface conditions appear to be as important as the pressure-recovery setting for organic rejection. Poor permeate quality during extended operation was primarily due to membrane fouling.

8) Inorganic parameters such as TDS and total hardness (TH) were reduced approximately 40% by UF at approximately 100 psi and 75% recovery.

9) The product water at VOG, AID, and Olga exhibited low DOC and color values that would not require additional color removal through chlorine bleaching.

10) The organic quality produced during the VOG and AID UF pilot plant tests were independent of the percentage of recovery and feed pressure for the test conditions.

11) The inorganic quality produced during the UF pilot plant tests at VOG and AID were independent of the feed pressures and improved as the percentage of recovery decreased.

12) UF with the N-50 membrane produced a stable, noncorrosive water.

13) The operation of the UF membrane pilot plant required less effort and skill than the operation of the other processes investigated, and it produced a vastly superior organic water quality.

2.4 Flotation

1) DAF of alum sludge caused floc shear post-precipitation, and it removed no more organics than conventional alum coagulation.

2) Foam flotation using conventional alcohols common to the mining industry removed no DOC from the raw water.

SECTION 3

RECOMMENDATIONS

1) PACl studies should be expanded to include high-turbidity and/or low-alkalinity waters.

2) DOC should be used as a surrogate for THM formation on a site-specific basis.

3) Specific organic rejection by membrane processes should be investigated to include priority pollutant removals and other organics capable of adverse health effects.

4) Cost and quality information on membrane processes should be made available to the field as soon as possible so that new plant expansion or construction can consider membranes and the corresponding water quality and cost.

5) A national survey of membrane plants should be conducted to determine (1) the cost of construction, operation, and maintenance, (2) - operating conditions and problems, (3) raw and finished water quality, and (4) brine disposal and membrane life.

6) Further bench and pilot plant testing using UF and RO membranes should be conducted on surface and ground waters of varying organic quality in order to ascertain the relationships among water quality, recovery, pressure, flux, membrane cleaning, membrane materials, and membrane life.

7) Newly constructed membrane plants using raw waters normally treated by lime softening or coagulation should be monitored for cost, water quality, and operating conditions, and that information should be published for field consumption.

8) A permanent membrane plant that uses UF and RO should be constructed, operated, and monitored for 2 to 5 years to provide for a small community (fewer than 3,000 residents) water that does not exceed the THM MCL's. Such a project would (1) establish membrane processes as a practical means of organic control, (2) provide the best possible data on cost, water quality, and operation, and (3) demonstrate the ease of operation, modular expandability, operator skill required, and consistency of membrane processes.

9) A mobile, pilot-plant, scale-hybrid system of low-pressure RO and UF membranes should be constructed and operated to assess organic control and production of waters with suitable hardness values. Also, saltwater-intruded, high-organic waters could be studied, particularly the coastal waters.

10) Research and development should continue on molecular weight rejection by membranes. This work has indicated that significant THM precursor rejection is accomplished between molecular weight rejection of 400 and 2000. A membrane with a pore size greater than 400 may have as good an organic rejection and be less expensive to operate.

SECTION 4

EXECUTIVE SUMMARY

4.0 General

The purpose of this cooperative agreement was to demonstrate the performance and cost of possible new technologies relative to conventional technologies of trihalomethane control. The scope of work was to identify new technology for THM control, identify sites where THM control was needed, and execute in-plant studies with the necessary cost documentation.

The technologies selected for this project were polyvalent aluminum chloride coagulation, sequential treatment of alum coagulation preceded by lime softening, membrane processes, and flotation. The selection of sites was coordinated with EPA Region IV and limited to Florida because the varying water supplies and THM problems in Florida allowed more technologies to be investigated within the project budget. Coordination with the Florida Departments of Environmental Regulation and Public Health resulted in the survey of more than 50 potential sites. Following plant visits, four sites were selected: (1) The Village of Golf Florida (VOG), a small lime-softening plant serving 2000 seasonal residents with THM's averaging in excess of 700 µg/L. (2) Acme Improvement District (AID), a lime-softening plant serving slightly less than 10,000 and averaging THM's of 380 µg/L. Both AID and VOG are near West Palm Beach, Florida, and utilize ground waters. (3) The Lee County Olga water treatment plant located near Fort Meyers, FL, is an alum plant which utilizes Caloosahatchee River water and had THM's of 700 µg/L. (4) Venice, Florida (the alternate site) serves 13,000 with a blended water from lime-softening and RO plants.

Optimization of existing plant processes for THM reduction was conducted at all four sites to develop comparable cost data and for the benefit of the plants. Bench and pilot plant membrane processes were investigated at VOG, AID, and Olga. Flotation experiments were conducted at Olga. Pilot plant sequential treatment was done at AID. Seasonal PACl/alum coagulation was studied on a plant scale at Olga.

4.1 Plant Optimization

Bench and plant scale investigations for plant optimization were conducted at three lime-softening plants located in the Village of Golf (VOG), Acme Improvement District (AID), and Venice, Florida. The bench tests clearly showed that the organic parameters studied (color, DOC, THMFP, and TOXFP) could be reduced by increasing softening pH. These plants typically softened at pH 10.3 to a total hardness (TH) of 100 ±50-mg/L as $CaCO_3$ and removed 20% to 30% of the raw color, DOC, THMFP, and TOXFP. The jar tests indicated an additional 1% to 2% DOC removal for every 0.1 pH unit increase in the reaction pH from 9.5 to 11.5. Generally, the THMFP reduction was greater than the DOC reduction because Cl_2 demand was decreased as a result of the lower DOC. The jar test studies also indicated 20-mg/L or less of alum would increase organic removal 5% to 10% during softening and would not be present in the finished water.

As a result of the bench scale studies, plant scale investigations were implemented at each of the lime-softening plants. Similar process changes were made at the three sites. These were (1) raising the reaction pH during softening from 10.3 to 11.0-11.3; (2) reducing the chlorine dose by eliminating chlorine as a means of color removal and pH reduction; (3) adding H_2SO_4 to offset the increased softening pH. At all three sites, THM's were reduced almost by half as were TOX and DOC relative to normal operations. Generally, 80% of the color was removed during softening and hardness was increased 50 to 80-mg/L due to the excess lime required to raise the reaction pH. THM's at AID and VOG averaged 240 μg/L and 330 μg/L during the plant test and were above the MCL. Venice reduced their THM's to 85 μg/L during the plant test and has adopted a high pH softening process for THM control. The additional cost of high pH softening ranged from \$0.02/1000 gal at Venice to \$0.22/1000 gal at VOG. The percent unit cost increase for high pH softening was 2% at Venice, 19% at AID, and 20% at VOG which corresponded to 32%, 50%, and 45% THM's reduction respectively.

4.2 PACl/Alum

The PACl/alum investigations were made on a bench scale and on a plant scale at the Olga water treatment plant located in Lee County, Florida, for both the wet and dry seasons. The organic parameters that were investigated were DOC, THM, TOX, and color as a result of coagulant dose and pH. Aluminum residuals, turbidity, sludge volumes and dry season softening were also investigated.

The raw water color averaged about 57 cpu during the dry season tests and increased 228% to 130 cpu during the wet season tests. Raw water DOC averaged 19.6-mg/L during the dry season plant tests and increased 9% to 21.4-mg/L during the wet season test. The raw turbidity average 6 NTU during the wet season and 4 NTU during the dry season. The alkalinity and hardness during the wet season were approximately 140 and 110-mg/L as $CaCO_3$, and increased to 250 and 160-mg/L as $CaCO_3$ during the dry season. The raw water used during both plant tests was a low turbidity, high color water that would be normally treated by sweep coagulation for color removal only. Two PACl coagulants, 100S- a sulfate chloride base and 190- a chloride base coagulant, plus alum were compared for organic removal.

Jar testing during the wet and dry seasons demonstrated the optimum coagulation pH was between 5.0 and 6.0 for DOC and color removal for all coagulants. PACl 100S was superior to PACl 190 for the removal of DOC and color in jar tests and was selected for plant testing. DOC removal increased with increasing coagulant dose between pH 5.0 and 6.0 but appeared to approach a limiting value. Coagulant doses of 117-mg/L as alum removed 60% of the DOC and 90% of the color in the wet and dry season jar test. Increasing coagulant dose to 175-mg/L increased DOC removal 5% to 10% and had less effect on color removal. The percent color and DOC removal during the wet season were slightly higher (>5%) than during the dry season for all coagulants. However the actual DOC concentration and, to a lesser extent, the color remaining in each season were approximately the same at equivalent conditions for each coagulant. Wet season THMFP reduction was very similar to DOC removal. The optimum pH range for THMFP reduction was 5.0 to 6.0 for all coagulants and increasing coagulant dose increased THMFP reduction in all jar tests. Doses of 117 to 175-mg/L as $Al_2(SO_4)_3 \cdot 14H_2O$ realized 90% of the maximum THMFP reduction during coagulation.

In all the jar testing alum generally removed slightly more (<5%) DOC, color, and THMFP than PACl. The seasonal jar tests did not indicate a major difference in coagulant demand for organic removal. This may have been due to the small DOC increase during the wet season. Aluminum residuals after coagulation with either PACl-100S or alum were typically less than 0.4-mg/L Al and were controlled by coagulation pH. These levels would not cause post precipitation problems. Removal of DOC and THMFP during coagulation should be in excess of 60% accompanied by more than 90% color removal. Although DOC, THMFP and color removal vary directly with equivalent coagulation dose and pH, DOC is the better indicator of THMFP than color.

Plant tests using alum and PACl were conducted in the dry and wet season for approximately 30 days each. The aluminum residuals in the distribution system were generally less than 0.2-mg/L. The aluminum residuals in the distribution system under normal plant process of using alum in conjunction with softening resulted in residuals of 5-mg/L Al. The seasonal plant data demonstrated finished Al residuals are controlled by coagulation pH which should be between 5.5 to 6.5 for minimum Al residuals. There was no significant difference between the sludge volumes produced during alum or PACl coagulation during the plant tests for equal operating conditions. The turbidity carryover from the settled water using either alum or PACl coagulation was typically less than 2 NTU and exhibited no differences between the two coagulants. A PACl 100S dose of 80-mg/L was not sufficient to produce a settleable floc during the plant test. An equal alum dose produced a settleable floc.

Softening preceded coagulation during the dry season plant test. Although softening removed 20% of the DOC and 43% of the color, no additional removal of color and DOC was realized by the succeeding alum or PACl coagulation than if coagulation was used alone. The addition of 1.2 meq/L of alkalinity to the softening process reduced the calcium hardness to only 100-mg/L as $CaCO_3$ although the alkalinity and total hardness were balanced. The softening pH was varied from 9.5 to 10.6 with 10.0 achieving the lowest hardness.

During the wet season the DOC was reduced an average of 57% to 8.6-mg/L by PACl and an average of 58% to 7.0-mg/L by alum. The dry season

DOC removals averaged 48% with reductions 10.1-mg/L for alum and averaged 39% with reductions to 12.1-mg/L for PACl. This lower reduction during the dry season was due to a higher coagulation pH caused by a hard $CaCO_3$ scale in the reactor. The THM's at the plant tap during the wet season averaged 235 μg/L for PACl and 248 μg/L for alum. The dry season THM's at the plant tap averaged 204 μg/L for PACl and 175 μg/L for alum which represented a decrease of 13% to 30% in the dry season. However TOX for alum or PACl were 710 and 714 μg/L during the wet season and decreased to 421 and 380 μg/L during the dry season. Color removal was similar to DOC removal exhibiting removals of 91 to 94% for PACl and alum during the wet season and 78% to 89% for PACl and alum during the dry season.

The DOC:THM:TOX ratios during the wet season averaged 1:32:94 and 1:16:39 during the dry season. The THM:TOX ratio did not vary as much as the DOC:THM:TOX ratio since both THM:TOX are directly effected by Cl_2 dose. The THM:TOX ratio was 1:2.9 and 1:2.4 during the wet and dry seasons, respectively.

The operation cost increased from $1.07/1000 gal to $1.16/1000 gal when the maximum alum dose was used during the wet season. This was an 8% cost increase which decreased plant THM's by 72%. The operational cost during the wet season for PACl was $1.62/1000 gal and similar THM's reductions were obtained. The cost of producing water was less during the dry season due to relatively constant labor and power costs and increased production. The plant costs were increased from $0.60/1000 gal to $0.78/1000 gal using the maximum alum dose. This was a cost increase of 23% and reduced THM's by 39%. Dry season PACl costs were $1.02/1000 gal and decreased THM's by 29%.

4.3 Sequential Treatment

Lime softening followed by alum coagulation, was investigated on a bench and pilot plant scale at the Acme Improvement District water treatment plant located near West Palm Beach, Florida. Jar tests were conducted to determine the best sequence of softening and coagulation for THM precursor removal. Pilot plant testing was executed in two unused reactors that were serially connected for the project. There was no turbidity in the raw water. The hardness and alkalinity were typically 350-mg/L as $CaCO_3$ with color and DOC values of approximately 40 cpu and 15-mg/L as C.

The initial bench scale sequential treatment work indicated that 60% of the DOC could be removed by sequentially treating AID raw water with softening and coagulation regardless of order. However, if pH was adjusted to 5.5 during coagulation, this DOC removal was increased 25%. Color removal was 85 to 90% and varied directly with DOC removal and THM precursor removal. THMFP was reduced to 167 to 217 μg/L by softening at pH 11.0 and coagulation at pH 5.5, which represented an increased average reduction of 45% relative to softening only at pH 10.3. TOX formation was generally three times THM formation.

The sequential treatment isopleths showed that softening at pH 11.0 averaged 41% DOC, 67% THMFP, 65% TOXFP and 63% color removal. This increased the removal relative to softening at pH 10.3 for each parameter (+11% DOC, +3% THMFP, +8% TOXFP, +3% color). Following lime-softening with alum coagulation also increased organic removal. The optimum pH and dose were 5.0 to 6.0 and 117-mg/L alum. The DOC, THMFP, TOXFP and color were reduced an additional 30%, 13%, 23% and 30% respectively. Alum coagulation following softening at pH 10.3 increased the percentage removal of all the organic parameters more than alum coagulation following softening at pH 11. However the lowest concentrations of DOC, THMFP, TOXFP and color were from samples softened at pH 11.0 and coagulated at pH 5.0 to 6.0 although these differences were typically less than 5%.

Plant testing of sequential treatment was conducted by softening at pH 10.3 and 11.0 followed by alum coagulation from pH 7.0 to 5.0. The alum dose was varied from 60 to 109-mg/L. The results show, as did the jar tests, that adding alum coagulation following softening will decrease the DOC, THMFP, TOXFP, and color relative to an existing softening process alone. The maximum DOC and color removals obtained in the plants tests were approximately 70% for DOC and 93% for color. The THMFP and TOXFP were reduced from 20 to 25%. There was slightly better removal for all parameters when pH 11 softening preceded coagulation as opposed to pH 10.3 softening. The organic removal achieved in sequential treatment indicates that DOC, THMFP, and TOXFP can be reduced an additional 20-25% when alum coagulation is utilized in series with softening. The chemical cost increase would vary from $0.10 to $0.15 per 1000 gallons which would increase AID O&M cost ($0.35/1000 gal) by 30% to 43%. However, adding softening to an existing coagulation process would not remove any additional organics.

4.4 Membranes

Bench (1000 gpd) and pilot scale (25,000 gpd) investigations of membrane processes were conducted at two sites, VOG and AID, that utilized ground water supplies. Bench scale investigations were also conducted at one site, Olga, that utilizes a surface supply. Initially, one RO and six UF (low pressure) membranes were purchased and tested for product (permeate) water quality on a bench-scale level. The UF membranes are designed for operation up to 100 psi, whereas the RO membrane is intended to operate at 200-250 psi. Bench results from all three sites demonstrated the only membranes capable of producing a water from these highly organic sources that would meet the THM MCL and maintain a Cl_2 residual was the Filmtec (UF membrane, N-50) and the RO membrane (BW 3030).

The nominal molecular weights rejected by each membrane were supplied by each manufacturer and ranged from 40,000 to 100. Bench scale testing indicated that a molecular weight rejection of 2,000 would typically pass 50% of the raw DOC but only 20% of the color. The resulting product THMFP was generally 800 µg/L for the surface water source and 400 µg/L for either ground water source. The UF with molecular weight rejection of 400 passed less than 10% of the DOC and 3% of the color at any site, while typically producing a water with a THMFP of less than 50 µg/L at the ground water sites and approximately 100 µg/L for the surface water. The product was essentially colorless and the DOC was 2-mg/L or less at all three sites.

The color, DOC, and THMFP of the RO membrane were approximately equal to the color, DOC, and THMFP of the best UF membrane. However, the RO membrane operated at 200 psi and rejected species of MW 100 or greater as opposed to the UF membrane which rejected species of MW 400 or greater at a pressure of 100 psi. The RO membrane rejected a much higher inorganic fraction than did the UF membrane. The RO membrane rejected over 90% of the TDS, total hardness (TH), Cl^-, and Na^+ at all sites, whereas the UF membrane inorganic rejection varied from 50% to 70% for the various parameters. Since the N-50 was the membrane that operated at the lowest pressure and still produced a water that met the THM MCL it was selected for extended operation.

The N-50 UF was installed at Olga in a bench scale unit capable of producing 1000 gpd and operated for 740 hours over a 45-day period. An operational percentage of recovery and feed pressure matrix was developed to determine the extended study operating conditions. Over operating conditions of 60 to 120 psi and 60% to 90% recovery the resulting product water quality was best at high pressure exceeding 105 psi and recovery of 60%. The matrix results indicated the THM MCL could be met if the operation conditions were 105 psi with 60% recovery. At these conditions product water quality and percent rejections were 1.6-mg/L DOC as C (92%), 3 cpu color (93%), 172-mg/L TDS (64%), 78-mg/L as $CaCO_3$ TH (65%), 60-mg/L Cl^- (40%), and 68-mg/L as $CaCO_3$ Alk (59%) with pH 7.8. During the extended operation, pressure was varied from 60 to 100 psi with percent recoveries of 60 to 90. The THM MCL was met for 105 psi, 60% recovery and for 75 psi, 60% recovery immediately after the membrane was chemically cleaned. The flux decreased with time during the extended study from 18 to 14 gpd/ft² over 150 hours of operation. After cleaning with the pressure at 95 psi, the flux declined from 22 to 16 gpd/ft² but the product quality remained constant. Flux was independent of recovery during the extended study. The water quality and percent rejection (%R) for the conditions meeting the THM MCL were 2.7-mg/L as C DOC (88%), 156 μg/L TOXFP (80%), and 3 cpu color (98%). The inorganic water quality was 145-mg/L TDS (65%), 68-mg/L Cl^- (32%), 85-mg/L as $CaCO_3$ TH (64%), and 7-mg/L as $CaCO_3$ Alk (58%). The pH was 7.8 and the water was stable.

A 25,000 gpd mobile UF pilot plant utilizing the N-50 membrane was built in a 30' trailer. The U.F. plant was housed in the 20x8' rear section of the trailer and equipped for antiscalant feed, acid feed, chlorination, stabilization, prefiltration, and storage as well as UF with variable recovery (50% to 90%) and feed pressure (80-120 psi). This plant was installed and operated at VOG for 365 hours from January 2 to March 3, 1985. Initially an operational test matrix was developed for water quality and flux from varying percentage of recovery and pressure. Product water DOC, color, and THMFP were independent of recovery and pressure over the test conditions and averaged less than 2-mg/L as C, 1 cpu, and 50 μg/L. Product water TDS and TH increased with increasing recovery, were independent of pressure, and varied from 25% to 75% of the raw water value. Operating conditions at VOG were set at 90 to 105 psi and 75% recovery in order to produce a water with a TH of 150-mg/L as $CaCO_3$, essentially no color, and THMFP of 50 μg/L or less. During the VOG operation, the product water quality and the percent rejection (% R) from the raw were 1.9-mg/L as

C DOC (88%), 3 cpu color (97%), 27 μg/L THMFP, and 47 μg/L as Cl TOXFP. The raw water, TDS, TH, and Alk were reduced to 195-mg/L (60%), 142-mg/L as $CaCO_3$ (62%), and 135-mg/L as $CaCO_3$ (60%), respectively. The final pH was 7.5 and the water had adequate alkalinity and hardness to be classified as stable which is contrary to RO products. The product water flux declined 32% during the VOG operation from 20 to 13.4 gpd/ft². The water temperature was approximately 25°C and essentially did not vary during the operation.

On March 3, 1985, the U.F. pilot plant was moved from VOG to AID and operated until May 31, 1985, with an elapsed time of operation of 1020 hours. A second operational test matrix was developed that showed color, DOC, and THMFP removal was independent of recovery and pressure over the test conditions (50% to 90% product recovery and 80 to 120 psi). Product water color, DOC, and THMFP were typically less than 2 cpu, 2-mg/L as C, and 30 μg/L. TH and TDS removal was independent of pressure and dependent on product recovery and varied from 25% to 75% of the raw water value. Operating conditions varied from 90-103 psi and 67-83% recovery. The product water quality and percent rejection (%R) from the raw were 1.5 cpu color (97%), 2.0-mg/L as C DOC (86%), 50 μg/L THMFP, and 48 μg/L as Cl TOXFP. The product water TDS, TH, and Alk values and percent rejection were 282-mg/L (43%), 187-mg/L as $CaCO_3$ (40%), and 180-mg/L as $CaCO_3$ (37%). The pH averaged 7.5 and the product was stable.

The UF pilot plant was designed with the membranes in four pressure vessels connected two each in series. The average pressure drop in the first pressure vessel was 32 psi (11 psi/membrane) and 41 psi (13 psi/membrane) in the second pressure vessel. The flux at AID rose slowly from 14 to 15 gpd/ft² during the first 150 hours of operation and after a chemical cleaning rose to slightly over 20 gpd/ft² and remained there for the duration of the study.

Cost of construction of a UF plant should be slightly less than an equivalent RO plant because of less expensive membranes and pressure requirements. Power costs should generally decrease from 75% (100 psi/400 psi) to 50% (100 psi/200 psi). The construction and O&M cost were estimated at $0.29/1000 gal and $0.47/1000 gal for a 1 MGD UF plant.

4.5 Flotation

Bench scale investigations of dissolved air flotation (DAF) in conjunction with and after alum coagulation of a high color, low turbidity water were conducted using a small DAF pilot plant supplied by Komline Sanderson. After contact with DAF, alum floc was found to shear, solubilize and post-precipitate after sampling was complete. DOC and color removal following DAF of alum coagulated waters were less than those having only alum coagulation. Foam flotation using alcohol based surfactants common to the mining industry did not remove DOC from the same high color, low turbidity potable water source.

4.6 Summary

The DOC removal from each of the processes investigated was different. Lime softening typically removed 10% to 30% of the raw DOC at normal reaction pHs of 9.0 to 10.3. The DOC removal during softening could be generally increased to 40% to 50%. Alum or PACl coagulation should remove 60 to 70% of the raw DOC if the reaction pH is from 5.0 to 6.5 and the coagulant dose is adequate, 80 to 180-mg/L as $Al_2(SO_4)_3 \cdot 14H_2O$ in the waters investigated. Sequentially treating water with lime softening and alum coagulation would remove 60% to 70% DOC and offers essentially the same removal as alum coagulation. RO at 200 psi or UF at 100 psi with selected membranes will generally remove 90% or more of the initial DOC. No other tested process other than membranes could achieve this high removal for a prolonged period. These general statements concerning DOC removals are supported by the test results of investigations conducted on a low turbidity high color surface water and low turbidity high color ground water.

Although DOC is a better surrogate than color for THM or TOX formation, DOC can not be used indiscriminately as a surrogate for chlorinated organic formation. These results and others have suggested that DOCs of less than 4-mg/L following conventional coagulation or softening are required to meet the THM MCL. Indeed, when the DOC was reduced to 5 to 6-mg/L by coagulation formation potentials of less than 200 μg/L were recorded. It was not possible to reduce the DOC any lower by coagulation or softening of these waters. However, DOCs of lower than 2-mg/L achieved by UF had formation potentials over 100 μg/L. The results then indicate the DOC is the best THM surrogate on a process basis. DOC as a whole from membrane processes are more reactive than DOC from precipitative processes. The THM reduction is significantly greater than the DOC removal because of the reduced chlorine demand which also reduces THM's. The plant experiences indicate that a 30% reduction in THM is possible if the process is tuned for increasing organic removal and minimizing chlorine dose.

CHAPTER 5

EXPERIMENTAL METHODS AND QUALITY CONTROL

5.0 Introduction

This chapter will present the experimental methods, laboratory procedures and associated quality control procedures employed for this project. Four different laboratory environments were involved. (1) Analyses were done in UCF environmental engineering laboratories on samples prepared at UCF and samples taken from the sites. (2) Analyses were done in the plant (site) laboratories by project and plant personnel. (3) Some analyses were done at commercial laboratories commonly used by the plant for testing beyond plant means. (4) Analyses were done at the Lee County Environmental laboratory, a State certified laboratory for drinking water testing to support all Olga work.

5.1 Analysis Description

Table 5 describes the procedure and tests used for each analysis. Quality Assurance procedures were followed for parameters tested on routine basis. These procedures included standard, replicate and spiked analyses as well as participation in EPA performance evaluations. Logs of all data collection and operation were kept at UCF and on site for each specific test or operation.

Only analyses which are not fully described in Standard Methods will be described here. Tests described in this section include: DOC, THM, THM's, Color, TOX, TOXFP and Aluminum.

Color

Color was determined by measuring absorbance at 460 nm. Water samples were filtered through a glass fiber filter (Whatman GF/A) and pH adjusted to 7.6 using concentrated acid or base as needed. The measured absorbance was compared with that of a standard curve of absorbance versus concentration for a series of known chloroplatinate solutions. Results from this method have been compared with the visual comparison method (Standard Methods 204.A). In this comparison (Taylor et al., 1984) results from the two methods agreed within 2 cpu for all samples.

Table 5. Summary of Analytical Procedures.

	Procedure	Description	Reference
1.	chlorine residuals	Amperometric titration DPD ferrous titrimetric	SM-408C SM-408D
2.	hardness, total Ca^{++} Mg^{++}	EDTA titration	SM-314B SM-311C SM-318C
3.	pH		SM-423
4.	Dissolved organic carbon (DOC)	oxidation to CO_2, reduction to CH_4, FID	SM-505
5.	inorganic carbon		SM-505
6.	Color	Absorption @ 460 nm	text
7.	Trihalomethanes (THM)	liq-liq extraction	text
8.	Trihalomethane Formation Potential (THM's)		text
9.	Total Organic Halogen (TOX)	Adsorption on GAC, to remove inorganics, combustion of sorbed TOX, microcoulometric silver ion titration of halides	text
10.	Total Organic Halogen Formation Potential (TOXFP)	Chlorination followed by TOX measurement	text
11.	Alkalinity	Acid titration	SM-403
12.	(TDS) Total Dissolved Solids	Evaporation at 103°C, gravimetric measurement	SM-209A
13.	Turbidity	Ratio Turbidimeter	SM-214A
14.	Aluminum (Al)	Plasma Emission	text
15.	Chlorides (Cl)	Argentometric titration	SM-407A

SM = 15th Edition of Standard Methods

DOC

Dissolved organic carbon was measured using a Dohrmann (Model DC-54) DOC Analyzer. This instrument operates by oxidizing organic carbon to CO_2 using exposure to UV light and acidic persulfate solution. CO_2 thus produced is catalytically reduced to methane and quantified using a flame ionization detector. Samples were collected in Pyrex glass bottles with teflon-lined caps, stored at 4°C, and analyzed within seven days of collection.

A DOC stock solution was prepared by dissolving 0.2125 g of potassium acid phthalate in 90 mL of reagent grade water, acidifying with 1 mL concentrated phosphoric acid and diluting to 100 mL. This stock solution contained 1000-mg/L organic C. The stock was stored at 4°C for no more than two months, when it was discarded and a new stock solution prepared.

To standardize the DOC analyzer, a 10-mg/L standard solution was prepared by diluting 1.00 mL of the stock solution to 100 mL using reagent grade water. This standard was analyzed and the instrument output adjusted to 10.000-mg/L plus the output of a blank solution. The detailed procedure for calibrating the DOC analyzer is described in the operating manual for the Dohrmann DC-54 DOC analyzer (Dohrmann DC-54 Manual, Santa Clara, CA).

Samples were filtered through a Whatman GF/A glass fiber filter and diluted with reagent water to a final concentration between 1 and 10-mg/L if the undiluted concentration was greater than 10-mg/L. Each sample was run a minimum of two times. If carryover from the previous sample was evident (i.e., the duplicate analyses differed by more than 10%) the sample was run again until repeatable results were obtained. All results are reported as averages of replicate analysis after correction for any DOC in the dilution water and carryover effects. The correction for organic carbon in the dilution water was to subtract the value of the blank from the analyzed value. Carryover effects were recognizable when two sequential samples had sizeable differences in DOC value. In such a case, the first replicate analysis of the second sample was influenced by the preceding (first) sample. In such cases the first replicate value was discarded and a third or even fourth replicate analyzed to insure the precision of the subsequent replicate analysis.

Trihalomethanes

Trihalomethanes (THM's) were concentrated from aqueous solution by liquid-liqud extraction using hexane as the organic phase solvent (Supplement to the 15th Ed. of Standard Methods Selected Analytical Methods Approved and Cited By the United States Environmental Protection Agency). Ten mL of the sample to be analyzed were carefully measured into a Pyrex test tube containing 2 mL of nanograde quality hexane. The tube was closed with a one-inch square piece of aluminum foil and a screw cap lid. The mixture was vigorously shaken for 30 seconds and the two phases allowed to separate. Three uL of the organic phase was used for analysis. The operating parameters for the gas chromatographic system are given in Table 6. Chloroform ($CHCl_3$), bromodichloromethane ($CHCl_2Br$), dibromochloro-

Table 6. Gas Chromatograph System and Operating Parameters.

Model	Hewlett-Packard Model 5750
Detector	Ni^{63} Electron Capture
Detector Temp	210°C
Carrier	5% Methane in Argon
Carrier rate	40 mL/min
Purge	5% Methane in Argon
Purge rate	120 mL/min
Column	6 ft Pyrex 3% SP-1000 on Supelcoport 120/100
Column temp	80°C

methane ($CHClBr_2$) and bromoform ($CHBr_3$) were identified and quantified by comparison to standard solutions. Samples outside of the linear range were reanalyzed after dilution of a fresh sample. THM standard solutions were prepared by diluting a stock solution, containing 200-mg/L each of chloroform, bromodichloromethane, dibromochloromethane and bromoform in methanol, in reagent grade distilled water. The 200-mg/L stock solution was obtained in 1 mL ampules (part #4-8746) from Supelco Inc. (NY).

Trihalomethanes (THM's) are defined in the following manner in this text. Trihalomethane formation potentials (THMFP) are defined by EPA as the difference between instantaneous Trihalomethanes (InstTHM) and terminal Trihalomethanes (TermTHM). These terms were developed to describe THM's throughout actual distribution systems. THMFP is measured by the difference after chlorinating a sample a second time that was taken from a distribution system and subtracting the InstTHM's measured before the second chlorination from the TermTHM's measured after the second chlorination. For the bench and pilot scale work in this project, all THM's are designated as THMFP since InstTHM was zero in all cases. These samples were all chlorinated in the laboratory and these laboratory induced THM's represent the formation potential corresponding to the 0.2-mg/L Cl_2 residual. The plant THM's are referred to only as THM's since at Olga and Venice some slight rechlorination occurred in the distribution system and reporting of InstTHM's and TermTHM's would create unnecessary complications. The point of significance is what was the measured THM's for specified conditions (pH, time, temperature, Cl_2 dose and DOC) relative to the 0.10-mg/L MCL. So the THM's from the Venice, AID, and VOG plant optimization, and the Olga PACl/alum wet season dry season studies are noted as THM's. The THM's from all membrane, bench, and pilot studies at Olga, VOG, and AID; the flotation work at Olga and the bench and pilot sequential treatment studies at AID are referred to as THMFP. This approach allows the THM reduction achieved by removing precursors on a bench or pilot scale to be described as THMFP reduction which is more accurate than THM reduction since no actual THM's were removed by any process, only their precursors or their formation potential as described.

Trihalomethane Formation Potential (THM's)

THM's was measured by adding sufficient chlorine to water samples at pH 7.6 so that a detectable free residual was maintained for 48 hours (unless some other time period was specified). At VOG and AID the chlorinated water was adjusted to pH 8.0. After the incubation period, the chlorine residual remaining was neutralized by adding excess sodium thiosulfate. THM's in this work was identical to terminal THM's since there was never any THM (InstTHM) found in laboratory samples prior to chlorination.

A 250 mL to 500 mL aliquot of the sample was placed in a beaker and the appropriate volume of chlorine stock solution added to produce a dose of 20 mgCl/L. (Raw water samples were chlorinated with higher doses (40-60-mg/L), depending on the nature of the sample, in order to maintain a free residual throughout the incubation period.) Immediately after chlorine addition, the pH was adjusted to 7.6 (or 8.0) using concentrated NaOH or

H_2SO_4 as necessary. The chlorinated sample was then poured into a 125 mL amber Pyrex glass bottle and fitted with a teflon-lined cap. After the incubation period, usually 48 or 96 hours, 1 mL of 0.1N $Na_2S_2O_3$ was added to neutralize the chlorine. Samples were stored in the same bottle at 4°C until analyzed. Analysis was usually completed within two weeks.

Chlorine Dissipation Studies

Chlorine dissipation rates were determined by monitoring free chlorine residual over time. Water samples, usually 1 L or more, were taken from either plant flow or laboratory texts. The samples were chlorinated at various chlorine doses, using reagent grade calcium hypochlorite as the chlorine source, and the pH adjusted to approximately the calcium carbonate saturation pH. The samples were poured into Pyrex glass bottles that had been heated to 550°C to remove organics and sealed with teflon-lined caps. These were incubated at the specified temperature for time periods up to 5 days. Small samples were removed at various times for chlorine residual analysis. THM and TOX analyses were performed on samples incubated in separate bottles, so they would not be affected by head space. These studies were intended to provide a basis from which to estimate applicable chlorine doses during plant tests. It was recognized that the chlorine demand of samples incubated in clean glass bottles was probably less than the chlorine demand of the actual distribution system.

Sludge Volume

Aluminum and calcium carbonate sludge volume were measured by allowing a sample of the sludge to settle in an Imhoff cone for a specified period of time (5 minutes for calcium carbonate sludge and 10 minutes for alum sludge). The volume occupied by the sludge was determining visually at that time by noting the height at which a definite interface existed between the settled and clear zones. The sludge volume was expressed as a percent of the total volume of the settling cone. This procedure gave a crude estimate of overall sludge quantities.

TOX

Total Organic Halogen was measured using a Dohrmann DX-20 Total Organic Halide Analyzer System (Xertex Corp, Santa Clara, CA). This instrument operates by adsorbing halogenated organic compounds onto a granular activated carbon column. After eluting inorganic halides with a KNO_3 wash, the GAC column is placed into a quartz combustion tube in which the GAC along with any adsorbed organic compounds are combusted. The halogen atoms are converted to hydrogen halide gas which is purged from the combustion tube into a micro coulometric titration cell in which halide ions are precipitated as silver halide, and silver electrodes are oxidized to replace the precipitated silver ions. The instrument measures the charge used in silver oxidation and converts this measurement to a digital read out in ug of equivalent Cl^- ion. Dividing this reading by the sample volume gives concentration in μg/L. A standard DOC solution is prepared by dissolving 1.856 g of 2, 4, 6 - trichlorophenol in 100 mL methanol. A 1000 μg/L standard is prepared from this solution by dissolving 100 μg/L of the stock solution into a liter of distilled water. Standard solutions of other concentrations were prepared by using proportionately more or less of the stock solution.

A blank carbon column was carried through the analytical procedure to correct for any background effects during the analysis. This blank value was subtracted from each analytical result. Samples that resulted in the detection of more than 40 ug of Cl were diluted with carbon filtered distilled water and reanalyzed. Further details of the operation of the Dohrmann DX-20 can be found in the operation manual.

TOXFP

TOX formation potential was measured on sample chlorinated in the same manner as THM's. A separate bottle was used for THM's, TOXFP, and any replicate samples.

5.2 Membrane Studies

5.2.1 Bench-Scale Membrane Studies at AID and VOG

Prior to bench scale testing at AID, seven membranes were chosen for testing for possible pilot plant use. These membranes were chosen on the basis on previous literature, manufacturer-technical specifications and consultation with personnel in the membrane field. The seven chosen membranes were as follows:

a) Filmtec BW30-4040 - Reverse osmosis membrane with 100 MW cutoff.
b) Filmtec N-50 - Nanofilter membrane with 400 MW cutoff
c) Filmtec UFP-4040 - Ultrafiltration membrane with 10,000 MW cutoff.
d) DeSal U90-G10 - Ultrafiltration membrane with 2000 MW cutoff.
e) DeSal U90-G50 - Ultrafiltration membrane with 20,000 cutoff.
f) Osmontics - OSMO PT-2 - Ultrafiltration membrane with 20,000 MW cutoff.
g) Osmontics OSMO PT-4 - Ultrafiltration membrane with 40,000 MW cutoff.

Given the chosen membranes project personnel selected the Filmtec BW30 (BW30), the Filmtec N-50 (N-50), and the DeSal U90-G10 (U90-G10) as an "optimum" membrane testing group. From the stated molecular weight cutoff values, it was expected that these three membranes would give the best organic and inorganic parameter removals. The remaining membranes were placed in a "secondary" test group and will be discussed in a following paragraph.

Testing of the optimum group at AID commenced on July 25, 1984, and concluded August 5, 1984. Each membrane was operated at least twenty-four hours to flush residual chemical solutions from the membrane before sampling. Each membrane was operated at a pressure that was suggested by the manufacturer and the bench-scale unit manufacturer. The conditions were as follows:

a) BW30 - 190 psi and 60% product recovery
b) N-50 - 60 psi and 65% product recovery
c) U90-G10 - 100 psi and 80% product recovery

It was felt these conditions would give a representative product water from each membrane which then could be compared to the other test membranes.

A modified BTI Selectro 007 unit was utilized and is described in detail in the Olga membrane studies section. It should be noted that the extensive pre-filtering necessary at Olga was not necessary at the AID or VOG site. All sampling and analysis procedures have been previously described. The parameters analyzed included pH, DOC, color, alkalinity, calcium and total hardness, turbidity, THM's, and chlorine residual.

During the period of September 10, 1984 to September 22, 1984 the secondary membrane group was tested at the AID site. This testing was not done at VOG because the initial testing had shown what membrane was required to meet the THM MCL and project personnel were occupied with plant optimization to gather more organic data. Previous to these dates, the membranes had been flushed for a minimum of twenty-four hours by the AID site plant personnel. Each membrane was tested at a range of feed pressures and product recovery rates. These operating conditions are given in Table 7. The parameters analyzed included DOC, color, total hardness, THM's and chlorine residual. The sampling and analysis procedures have been previously discussed.

At the VOG sites the optimum membrane group was tested during the period of August 13, 1984 to August 16, 1984. Each membrane was flushed approximately six hours before sampling. The testing procedure and operating conditions, were identical to those found at the AID site. It should be noted that the secondary membrane group was not tested at the VOG site.

5.2.2 Bench Scale Membrane Testing at Olga

Ultrafiltration membrane testing at Olga was initiated on October 31, 1984, with an optimal membrane selection study to select the "best" membrane for organic and inorganic removal from the Caloosahatchee River source. The same seven membranes, having different molecular weight cutoffs, were utilized for testing. Each membrane was operated for a twenty-four hour acclimation period, before sample collection. Laboratory analysis, with the exception of organic parameters was conducted by the Lee County laboratory. Membrane selection studies lasted until the end of January due to operational problems and holiday delays. The major operational difficulty was the prefilter clogging from direct alum feed into the raw water source immediately before the raw water pumps. Details of the optimal membrane selection study are discussed in a later section. The optimal membrane selected was the Filmtec N-50 Ultrafiltration membrane.

In order to lessen membrane clogging due to high turbidity and direct alum feed, a prefiltration system was proposed. A shallow well on the river's edge was drilled, to be used as a "sand prefiltered" raw water source for the membrane unit. Organic sediments surrounding the well

Table 7. AID Secondary Membrane Testing-Operating Conditions

Membrane	Feed Pressure (psi)	Product Recovery Rate (%)
Filmtec UFP-4040	30	76
Filmtec UFP-4040	50	77
Filmtec UFP-4040	75	80
Filmtec UFP-4040	100	82
DeSal U9U-G50	35	72
DeSal U9U-G50	50	73
DeSal U9U-G50	75	77
DeSal U9U-G50	100	78
Osmontics OSMO PT-2	30	78
Osmontics OSMO PT-2	40	85
Osmontics OSMO PT-2	50	85
Osmontics OSMO PT-2	75	86
Osmontics OSMO PT-4	25	83
Osmontics OSMO PT-4	30	83
Osmontics OSMO PT-4	40	84
Osmontics OSMO PT-4	50	87

yielded flows less than 2 gallons per minute at low pressure. A series of modifications to the prefiltration system yielded a final configuration on April 11, 1985. During the system modification, two pressure-percentage of recovery matrices (February 28 and March 29) were conducted to determine optimal pressure-recovery operating conditions. Matrix results are presented as part of the N-50 Ultrafiltration Study.

On April 11, 1985, an extended operation study was initiated to define long term performance of the optimal membrane. The objective was to attain the optimal organic quality (low TOXFP, THM's) at a reasonable recovery rate. Samples were collected by Lee County or UCF personnel twice a week, for a forty-day period. Raw, permeate, and reject samples were taken for mass balances. A total of 16 sample sets were collected for organic and inorganic analysis. Laboratory analysis was largely conducted at UCF. Organic parameters investigated were DOC, THM's, TOXFP and true color. Inorganic parameters which were measured were Alk, TH, CaH TDS, Chlorides, and Sodium. Sodium analysis was done by Lee County. On two occasions, sample sets were shared by Lee County and UCF for independent inorganics analysis.

In spite of the prefiltration system, hydraulic problems affected the performance of the pilot scale membrane unit, during the extended study. Membrane fouling caused by organic and inorganic components of the river source reduced permeate flow rates. Low permeate flows at lower feed pressures (<75 psi) resulted in very low reject flows to maintain a recovery greater than sixty percent. The reject valve tended to close over time at reduced flows causing higher recoveries. The results and discussion of the N-50 UF study is presented in the results section.

5.2.3 Selectro "007" Bench Scale Membrane Unit

The "007" bench scale unit was used for short term membrane selection testing at the ACME, Village of Golf and Olga Water Treatment Plants. It was also used for an extended operation study at Olga. A flow schematic showing the major components appears in Figure 1.

Raw water (from a ground or surface source) under pressure enters the prefilter cartridge vessel, where it can be combined with recirculated reject water. The combined stream is passed through 5-micron size woven fiber filter to remove suspended solids or colloidal material. The cartridge is the type commonly used for swimming pool filters. The combined stream of prefiltered raw and reject water enters a pressurizing pump, which raises pressure to the desired level for UF. Pump output pressure is controlled by an adjustable port which is partially closed to increase pressure. Flow through the Procon brand pump is relatively constant at 4 gpm over a 50 to 120 psi range.

Pump-pressurized water is fed into the membrane vessel and distributed over the end surface of the spiral wound membranes, which were used in this project. Feed water is "ultrafiltered" dividing the membrane influent into a "product" water stream which passes through the membrane, and a "reject" flow which passes across the membrane surface. Product water or "permeate"

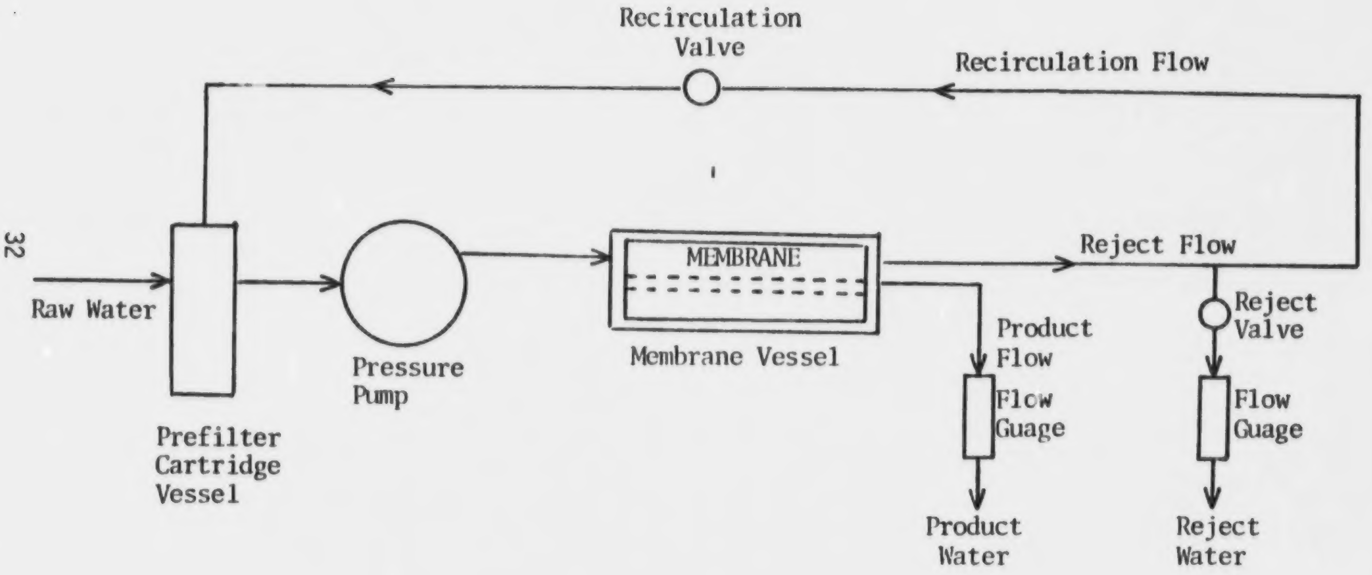

Recirculation Valve
Recirculation Flow
Raw Water
MEMBRANE
Reject Flow
Reject Valve
Prefilter Cartridge Vessel
Pressure Pump
Membrane Vessel
Product Flow
Flow Guage
Flow Guage
Product Water
Reject Water

Figure 1 Flow schematic of selectro "007" bench scale membrane unit. (Modified by Basic Technologies, Inc.)

is collected into a central collection tube and passed out of the membrane vessel. Product water flow is measured by a variable area flow meter (ball rotameter), and discharged to atmospheric pressure, through an outlet tube.

The reject flow in the reject pipe can be split (if desired or necessary) into two streams, a reject stream and a recirculation stream. The reject water is passed through a control valve, a ball float rotameter, and discharged to atmospheric pressure. The recirculation stream is routed back to the prefilter vessel through a recirculation valve.

The bench scale unit is designed to operate over a range of feed pressures (0-200 psi) and product flow conditions. Product flow depends on the feed pressure applied, the membrane type (MW size and material), condition of the membrane surface (clean or fouled) and the raw water source characteristics. "High" molecular weight cutoff membranes produce more product water than "tighter" membrane, for a given feed pressure and surface condition. A "clean" surface with open molecular "pores" will filter more water than a clogged membrane at a fixed feed pressure. A major operational constraint was the requirement of the constant speed pressurization pump to deliver about four gallons per minute, over the pressure range, irrespective of product and reject flow rates. Insufficient "feed" volume or pressure could cause cavitation and laboring of the pump. Consequently, the unit was equipped with a recirculation control valve to allow the operator to control recirculation flow. The amount of recirculation required to meet the pump limitation depended on the amount of water removed from the system in the product flow, and the desired recovery ratio (product flow divided by input raw water flow). The recovery ratio or percentage of recovery was adjusted by the reject flow valve.

The effects of product flow, percentage of recovery, and recirculation flow will be illustrated in the following examples.

1) Assume product water flow is 1.0 gpm at a 50% recovery ratio. Therefore 1.0 gpm is 50% of the raw input flow (2 gpm), making reject flow equal to 1.0 gpm. To satisfy the pump constraint, 4 gpm must pass through the pump. 2.0 gpm must be recirculated to make up the requirement. Thus 50% of the blended "feed" water contacting the membrane surface is recirculated "reject" water.

2) Assume product flow is very low (0.25 gpm) due to low feed pressure or a very low molecular weight cutoff membrane. Percentage of recovery is set for 50% of the raw water input. The reject flow would be 0.25 gpm, making the raw input equal to 0.5 gpm. The pumping limitation requires the recirculation flow rate to be 3.5 gpm. Therefore 87.5% (3.5/4.0) of the feed flow consists of recirculated "reject" water.

Obviously, the greater the product flow, the lower the amount of flow which must be recirculated for a set percentage of recovery. The amount of recirculated "reject" flow and the reject water quality largely determines the "feed" water quality, which ultimately affects the permeate quality. Water quality results from bench scale testing at AID, VOG, and Olga will illustrate the effect of product flow, and percentage of recovery on product water quality.

5.2.4 Membrane Pilot Plant Construction and Operation at AID and VOG

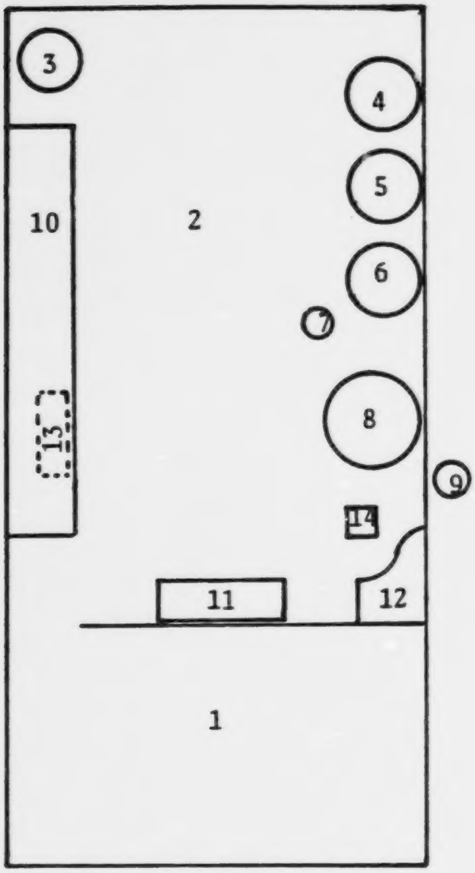
In order to further test the Filmtec N-50 Nanofiltration membrane, a plant unit capable of producing approximately 18,000 gpd of product water was utilized at the VOG and AID sites. The unit was constructed during the Fall of 1984 by Basic Technologies Inc. (BTI) of Riveria Beach, Florida to system criteria supplied by UCF. Specifications for the unit were supplied by the project personnel, the membrane manufacturer (Filmtec Corporation) and BTI. The unit utilized, in part, components from a RO unit previously used in an EPA project at Charlotte Harbor, FL.

The unit was designed to be reasonably self-contained and easily transportable between testing sites. It is shown in Figure A-1. To this end the components of the pilot unit were housed in a construction office trailer measuring 30 feet long and eight feet wide. A schematic of the trailer can be seen in Figure 2. The front portion of the trailer is an office that was used for sample analysis, data keeping and storage. The back section of the trailer houses the pilot plant components including the pressure vessels, control board, wet well, pumps, and chemical holding tanks which are all shown in Figures A-2, A-3, and A-4. Hydraulic connections are provided on one side of the trailer for raw, product and reject water lines. A single electrical cable provides all necessary power from a 220 volt source. Finally provisions are made to secure the aeration tower to the side of the trailer. The hookup and disassembly of the unit is easily accomplished in several hours by a three-man crew and the entire unit is transportable by a 1-1/2 ton pick-up truck.

A flow diagram is given in Figure 3. Raw water for the unit is supplied via 1-1/4" diameter lines from the existing treatment plant raw water lines. During the initial set-up of the unit it was found that the existing raw water pressure was not sufficient to feed the UF pumps. Subsequently a Sta-Rite P2R centrifugal pump was installed in-line at the raw water intake to provide adequate pressure.

After entry into the trailer chemical injection ports are available for acid and anti-scalant (Flocon-100) addition. During the testing period acid addition was found unnecessary, while 5-mg/L of Flocon-100 was applied on a continuous basis. A temperature probe is also provided in this section for raw water temperature monitoring.

Subsequent to this section, the raw water line is joined with the concentrate recirculation line. At this point the stream is known as feed water. The feed water is then prefiltered using a Filter Specialists, Inc. Model FS-85 Bag Filter with a V-rayon-viscose felt 10u filter bags. During the VOG site testing a previously used stainless steel prefilter with 10u filter cartridges was used. Due to a persistent leakage problem around the top, this filter was discarded.



1. Office Area
2. Membrane Pilot Plant
3. Acid Storage Tank w/ Pumps
4. Anti-Scalent Tank w/ Level Switches, Pumps and Mixers
5. HTH (Chlorine) Tank w/ Level Switches, Pumps and Mixers
6. Soda Ash Tank w/ Level Switches, Pumps and Mixers
7. Pre-filter
8. Clear Well w/ Level Switches
9. Degasifier
10. Membrane Pressure Vessels and Instrument Panels
11. Control Panel
12. Counter w/ Sink
13. Tonkaflo Pump
14. Transfer Pump

Figure 2 Membrane Pilot Plant Trailer Schematic

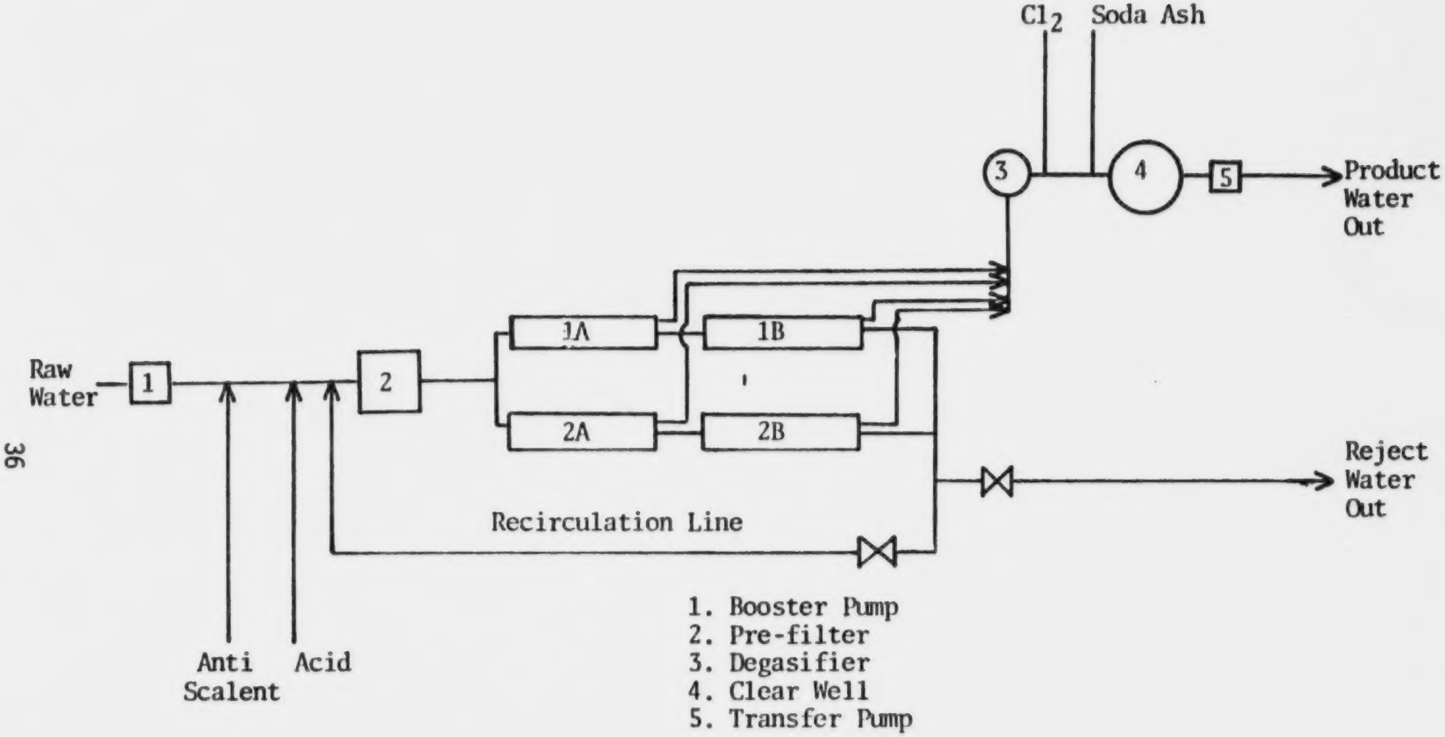

Cl2
Soda Ash
Product Water Out
Raw Water
1
2
3
4
5
1A
1B
2A
2B
Reject Water Out
Recirculation Line
Anti Scalent
Acid
1. Booster Pump
2. Pre-filter
3. Degasifier
4. Clear Well
5. Transfer Pump

Figure 3 Membrane Pilot Plant Trailer Flow Diagram.

From the prefilter the feed stream is directed to a Tonkaflo 2008 AZ-3080 pump which provides the necessary pressure for the membrane process. The pressurized feed flow is delivered to a manifold where the flow is split and directed to the feed end of the pressure vessels 1A and 2A. Each pressure vessel contains three spiral wound membranes in series. The permeate, or product water is collected in a common center tube and, at the end of the pressure vessels, is routed to the product water collection manifold. Product flow is regulated using a flow restrictor prior to the manifold. The concentrate, or reject water from IA and 2A becomes the feed water for the pressure vessels 1B and 2B respectively. Following these pressure vessels (again each containing three membranes in series) the permeate is routed to the product water collection manifold while the concentrate is collected in the reject water manifold.

The product water is then routed to the top of the aeration tower. Hydrogen sulfide is stripped through counter-current aeration. The product water is collected at the bottom of the aerator and routed back into the trailer. Prior to the clear well injectors are available for chlorine and sodium carbonate (alkalinity) addition. From the clear well the product water is routed out of the trailer using a Sta-Rite P2R centrifugal pump as a transfer pump. At both locations the product water and reject water were combined and used in the existing treatment plant as raw water.

The reject water, after collection in the reject water manifold, is routed to a junction point. At this point desired percentages of the concentrate can be shunted to either the reject line or the recirculation line using valves on each line. An increase in recirculation at a given flow rate will have two effects: the first is the increase the percent product recovery rate and, secondly, increase the feed water quality parameter concentration.

Flow and pressure gauges are available within the system. Flow is measured on the combined product water stream, the recirculation stream and the reject stream. These three flows allow the calculation of the following parameters:

a) Raw Water Flow - the sum of product water and reject water.
b) Percent Product Water Recovery - the result of dividing the product water flow by the raw water flow and multiplying by 100.
c) Percent Recirculation Rate - the result of dividing the recirculation flow number by the sum of the recirculation flow and raw flow and multiplying by 100.

Pressure can be measured in several areas of the system. Pressure is measured on the feed flow at the prefilter inlet and on the suction side of the Tonkaflo pump. Feed pressure is measured in the feed water manifold. Concentrate pressure is measured in the reject water manifold. Throughout a good deal of the testing the product water pressure was measured at the product water collection manifold. However, due to the pressure-reducing capabilities of the flow regulators, this value was found to reflect only the back pressure from the aeration tower. Thus, during the testing at AID, this pressure reading was discontinued and the system was modified to obtain pressure readings on the four product water streams (from 1A, 2A, 1B, and 2B) prior to the flow regulators. The measurement of the above pressures are used to calculate the following parameters:

a) Prefilter Loss - the difference between the prefilter inlet pressure and the Tonkaflo pump suction pressure. Gives indication of prefilter bag fouling.
b) Transmembranic - the difference between feed pressure and the individual product flows from pressure vessels 1A, 2A, 1B and 2B. Two points should be noted in that first, since each pressure vessels holds three membranes, the pressure loss is over the three membranes. Secondly the feed pressure to pressure vessels 1B and 2B is less than the initial feed pressure. Measurements have shown this difference in feed pressure to be 9 psi on a constant basis.

As membranes become fouled, cleaning is necessary to restore lost flux. Cleaning of the membranes in the pilot unit is possible using the existing clear well and transfer pump. A hose connection is available to the feed water manifold from the transfer pump with return hoses from the reject and product water lines to the clear well. The clear well is used for chemical cleaner mixing. The cleaning procedure is as follows:

a) A 0.1% phosphoric acid solution is circulated for thirty minutes.
b) A combined basic solution of 1% Sodium EDTA, 1% Trisodium Polyphosphate (TSP) and a 1% Sodium Tripolyphosphate (STP) is circulated at a temperature of 40°C for one hour.
c) The membranes are rinsed with raw water for one hour with periodic venting of the rinsing solution to waste.

Aside from the major cleaning described above, a flushing of the membranes was done periodically. This involved opening the reject valve to allow maximum reject water flow (no recirculation). This action increases the flow velocity across the membrane surface and help "scour" particles from the surface. While the ability of this flushing to improve flux is questionable, such action undoubtedly reduces pressure losses through the feed and brine channels in the membranes.

5.2.5 Membrane Pilot Plant - Long-Term Testing at AID and VOG

Initial start-up of the unit was attempted on 12/13/84 at the VOG. The necessity of adding a booster pump and the repair of several leaks delayed the start-up to 12/20/84. On 12/27/84 the staff at VOG were trained in the operation of the pilot plant with long term testing beginning 1/2/85. The pilot plant was operated on a daily schedule that corresponded to the existing treatment plant schedule; in this case approximately six hours a day. The unit was operated continuously from 1/2/85 to 3/3/85 with two days of non-operation (2/1/85 and 2/3/85). The unit was operated for sixty days for a total elapsed time of 365 hours. Pressure, flow, and time of operation data were recorded daily. Raw, reject and product water samples were collected every three days (approximately every 18 hours). Beginning on 2/2/85 a feed water sample was also collected in conjunction with the other samples. The following water quality parameters were analyzed for each sample; Dissolved Organic Carbon (DOC), True Color, Alkalinity, Total Hardness (TH), Total Dissolved Solids (TDS) and pH. In addition product water samples were chlorinated in the lab for Trihalomethane and Total Organic Halide formation potential. The samples were chlorinated at doses of 7-mg/L as Cl_2, adjusted to pH 8.0 and held for 96 hours at 25°C.

A major cleaning of the membranes did not occur during the VOG site testing. However system flushes were performed on 1/18/85 (106 hours elapsed time), and 2/2/85 (190 hours elapsed time). In both cases the system was flushed at a high reject water rate (> 18 gpm) for at least 30 minutes.

Initial conditions for operation (1/2/85) were set at a feed pressure of 90 psi and a product recovery rate of 75%. Initial product flux was 20.0 gallons/day-ft^2 (12.5 gpm). These pressure and recovery rate conditions were reset on 1/18/85 but the product flux had deteriorated to 18.4 gpd/ft^2. Flushing partial restored the flux to 18.7 gpd/ft^2. On 2/2/85 after approximately 190 hours of operation the operating conditions were changed to a feed pressure of 100 psi and the product recovery rate reset to 75%.

In addition to the long-term testing, the pilot unit was utilized on 1/18/85 and 2/2/85 for short-term tests involving varying feed pressures and product recovery rates. On 2/2/85 a 4x4 testing matrix was run using feed pressures of 80 psi, 90 psi, 100 psi and 110 psi and product recovery rates of 50%, 66%, 75% and 90%. In this manner sixteen combinations of feed pressure and product recovery rate were obtained. The product waters were analyzed for TDS, TH, THM's and true color. Raw water characteristics were determined by the analysis of three periodic raw water samples during the daily test.

On 1/18/85 testing was done on product waters obtained at a constant feed pressure of 90 psi and product recovery rates of 50%, 66%, 75% and 90% both before and after a hour-long membrane flush. In addition samples were obtained at a constant product recovery rate (75%) and varying feed pressures (80 psi, 100 psi, and 110 psi). In addition separate product water stream samples were obtained at 90 psi feed pressure and a 75% product recovery rate.

Subsequent to shutdown at VOG on 3/3/85, the pilot unit was transferred to AID on 3/7/85. Prior to start-up on 3/12/85, the existing prefilter was changed and an electrical transformer to convert 440V power to 220V power was installed. On 3/12/85 and 3/13/85 a 4x4 test matrix was run using feed pressures of 80 psi, 90 psi, 100 psi and 110 psi and product recovery rates of 50%, 60%, 75% and 90%. The product samples were analyzed for DOC, TH, THM's, TDS and true color. Raw water characteristics were determined from averaging several periodic raw samples.

Long-term testing commenced on 3/14/85 and terminated on 5/31/85. Again the pilot unit operated on a schedule that corresponded to the treatment plant with a daily operating average of 12.5 hours. The unit operated for 81 continuous days. On 5/20/85 the unit operated for less than one hour due to severe electrical storms. A total elapsed operating time of 1018 hours was observed.

The sampling schedule was similar to that at VOG with samples taken every 3 days (approximately 37 hours). Every sixth day additional samples were taken of the separate product waters 1A, 2A, 1B, and 2B. Pressure, flow, temperature and time of operation data were taken daily.

Initial conditions for operation were set at a fed pressure of 90 psi and a product recovery rate of 70%. Initial flux was 13.6 gpd/ft² (8.5 gpm). On 3/26/85 the feed pressure was reset at 90 psi. On 4/9/85 (367 hours elapsed time) the feed pressure was set at 100 psi and the product recovery rate at 74%.

On 5/6/85 the pilot plant membranes received a major cleaning using the procedures described previously. Flux had actually improved steadily during the AID testing rising from 13.6 gpd/ft² to 14.4 gpd/ft² prior to the increasing of the feed pressure to 100 psi and to 15.4 gpd/ft² subsequent to this action. However an approximate 23% flux loss had been noted since initial start-up at VOG. Subsequent to cleaning flux improved to 20.8 gpd/ft² which slightly exceeds the initial flux value. It should be noted that the initial flux was measured at 90 psi while the post cleaning flux was measured at 100 psi.

5.3 Jar Tests

Jar tests were conducted using raw water from the various test sites. Raw water samples were collected in 20 L polyethylene containers and transported immediately back to the UCF laboratory. These samples were stored at 4°C until the jar tests were performed. At that time, the samples were allowed to come to room temperature before performing the jar tests.

Jar tests were conducted in 1.0 L containers using a standard six place Phipps and Bird jar test machine. Coagulants and coagulant aids were added during a 1 minute rapid mix with a mixing speed of 100 rpm. Sulfuric acid or sodium hydroxide for pH adjustment were added during the rapid mix and slow mix period as necessary to maintain the desired pH. The mixing speed during the slow mix period was reduced to 20 rpm. The slow mix period lasted twenty minutes. Following slow mix, the sample was allowed to settle for 20 minutes, before filtering through a glass fiber (Whatman GF/A) filter.

Laboratory jar tests were performed using raw water from the Olga plant taken at times representing both the wet and dry seasons. The purpose of these jar tests was to determine the effect of coagulation dose and coagulation pH on the performance of alum and PACl.

The values of coagulant dose and pH chosen for the jar tests are shown in Table 8 and 9. The wet season coagulation doses are higher than the dry season doses due to the increased organics in the raw water. For the dry season tests, each coagulant was used at a dose of between 29 and 175-mg/L

Table 8. Experimental Matrix for Dry Season Jar Tests
Olga Plant Demonstration

				pH		
Coagulant	Dose	4.5	5.0	5.5	6.0	7.0
Alum	29	x	x	x	x	x
	58	x	x	x	x	x
	116	x	x	x	x	x
	175	x	x	x	x	x
PACl	29	x	x	x	x	x
	58	x	x	x	x	x
	116	x	x	x	x	x
	175	x	x	x	x	x

Table 9. Experimental Matrix for Wet Season Jar Tests
Olga Plant Demonstration

				pH		
Coagulant	Dose	4.5	5.0	5.5	6.0	7.0
Alum	29	x	x	x	x	x
	58	x	x	x	x	x
	116	x	x	x	x	x
	175	x	x	x	x	x
	191	x	x	x	x	x
PACl	29	x	x	x	x	x
	58	x	x	x	x	x
	116	x	x	x	x	x
	175	x	x	x	x	x
	191	x	x	x	x	x

as alum. PACl and alum dose were first expressed as Al_2O_3 but are expressed as the equivalent alum ($Al_2(SO_4)_3 \cdot 14H_2O$) dose for reader convenience. The pH in these tests was varied between 4.5 and 7.0 using sulfuric acid or sodium hydroxide as needed. The coagulant and 0.5-mg/L of polymer (Calgon 243) were added during the rapid mix along with the required acid or base to result in the desired pH. The same basic procedure was followed for the wet season jar tests, except that more values of pH (6.5 and 8.0) and coagulant dose (291-mg/L) were included. Following filtration dry season jar tests samples were analyzed for DOC and color to determine the optimum coagulation pH for these parameters. Jar tests at this pH were then repeated and THM, TOX and aluminum analyses performed in addition to DOC and color. In the wet season jar tests, DOC, THM, Al, and color were analyzed on all samples and TOX was analyzed for the pH 5.5 samples.

5.4 Aluminum Coagulants

The aluminum coagulants used in this study were alum ($Al_2(SO_4)_3.14H_2O$) and polyhydroxyaluminum chloride (PACl). The properties of these two coagulants differ as a result of differing manufacturing conditions. Alum is manufactured by treating aluminum bearing ores with sulfuric acid and then purifying and concentrating the resulting aluminum sulfate solution. PACl, on the other hand, is manufactured using a proprietary process by the addition of base to a partially hydrolyzed aluminum salt solution. Sulfate may be excluded from this process. Of the two PACl products tested, one excluded sulfate (AC 190) and the other included sulfate (AC 100S). These are noted in the text as PACl 100S and PACl 190. Differences between the two PACl products and alum presumably result from the differences in manufacture that result in the formation of a polymerized structure of PACl, and the difference in sulfate content for the three products.

Polyhydroxyaluminum (PACl) was supplied by PPG Industries Inc. (Pittsburgh, PA) in two different products, Prodefloc AC 100S and Prodefloc AC 190. Specifications for these products are given below:

	Prodefloc AC 100S	Prodefloc AC 190
Appearance	Pale Yellow Liquid	Amber Liquid
Specific Gravity	1.12-1.28	1.32-1.42
Viscosity @ 20°C	4.3-5.2 cp	19.7 ±2.7 cp
Freezing Point	-10°C	-15°C
pH @ 20°C	2.0-2.6	1.4-1.8
Al_2O_3 content	9.7-10.3%	17-19%

Alum ($A_2(SO_4)_3 \cdot 14H_2O$) was provided in either a liquid or powdered form by Allied Chemical Co. Specifications for liquid and powdered alum are given below:

	Liquid Alum
Appearance	Light Yellow-Green Liquid
Specific Gravity (60°F)	1.335
Viscosity (60°F)	26 cp
Freezing Point	-15°C
pH (1% solution)	3.5
Al_2O_3 content	8.3%

	Powdered Alum
Appearance	White to cream color
Bulk Density (lb per ft3)	38-45
Typical screen size	97% through 100 mesh
pH of 1% solution	3.5
Al_2O_3 content	17%

5.5 Quality Control

During the initial phases of the project, various measures were taken to insure analytical precision and accuracy. These measures include strict adherence to analytical procedures described above, careful cleaning of all glassware, proper collection and storage of samples, regular maintenance of analytical equipment and data collection for quality control purposes. This data gathering included the analysis of replicate samples, analysis of spiked samples, and participation in EPA Performance Evaluations. Initial data were utilized to establish quality control limits for continuous monitoring of analytical precision and accuracy. Subsequent quality control data compared to these limits included the analysis of spikes, replicates and participation in EPA Performance Evaluations.

5.5.1 Precision and Accuracy

DOC precision was determined by running a series of duplicate analyses on samples throughout the study period. The results of these analyses were used to calculate the industrial statistic, I, as described in the EPA Handbook for Quality Control in Water and Wastewater Laboratories.

$$I = |A-B|/(A+B)$$

The average and standard deviation of this statistic were used to define the acceptable region for duplicate analyses. The upper control level (UCL) was established at the average value of I plus 3 standard deviations. The I values above the upper control level were deemed out of control and all analyses halted until the cause of the out of control condition was ascertained and remedied. Values of I that exceeded the average by more than two standard deviations were considered to be in the upper warning region (UWL). If values occurred in this region, analytical procedures were reviewed for possible errors, but analyses were not halted.

Results of duplicate analyses for DOC are given in Table 10 along with calculated values for I. The average value for I was 0.0318, indicating that the average variability of DOC analysis was 3.2 percent. The upper control level was 0.1327 and the upper warning level 0.0991. Table 10

Table 10. Industrial Statistic for Duplicate DOC Analyses

Duplicate		
A (mg/L as C)	B (mg/L as C)	I
9.76	10.15	0.0195
9.24	9.76	0.0273
22.61	18.04	0.1124
8.07	7.24	0.0542
11.32	11.82	0.0216
7.68	6.91	0.0527
21.34	22.22	0.0202
13.86	14.38	0.0184
8.29	8.20	0.0054
15.99	16.94	0.0288
19.04	17.65	0.0378
18.43	18.98	0.0147
19.16	18.72	0.0116
19.05	18.54	0.0135
9.60	9.54	0.0031
9.81	8.53	0.0697
4.27	4.57	0.0339
18.17	18.76	0.0159
0.64	0.47	0.1531*
14.6b	15.23	0.0190
4.32	4.24	0.0093
1.36	1.26	0.0381
0.96	0.97	0.0051
1.76	1.56	0.0602
1.90	1.81	0.0242
13.84	16.33	0.0825
1.95	1.95	0.0000
12.55	13.20	0.0252
2.27	2.33	0.0130
4.24	4.07	0.0204
2.39	2.40	0.0020
3.14	3.17	0.0047

AVG. = 0.0318
s = 0.0336
UCL = 0.1327
UWL = 0.0991

*Exceeds UCL

shows that one sample in the 32 duplicates exceeded the upper control level. This sample had an average DOC concentration of 0.56-mg/L which was relatively low compared to the other samples. The analyses performed on the same day as this duplicate were repeated and subsequent duplicates were in control. Poor reproducibility in duplicate analysis was usually traced to a gas leak in the DOC analyzer. Disassembling, cleaning, and reassembling the glass spargers was usually sufficient to restore the instrument to proper operation.

The DC-54 was calibrated at the start of each sample run using a 10 ppm DOC standard. This standard was replaced every two months to insure its concentration had not decayed. Previous analysis of EPA Performance Evaluation Standards indicate that percentage of recovery of standard solution is 100%, (Table 11).

Data for 36 sets of duplicate TOX analyses are given in Table 12. These data represent duplicate analyses of both standard and unknown samples. The average of I was 0.028 and the UCL was 0.081. The UWL of 0.075 was exceeded in one analysis. Subsequent analyses were all acceptable, so no corrective action other than routine maintenance of the TOX analyzer was performed. In addition to these duplicate analyses,multiple replicate analyses were performed on several standard samples of known concentration. These values are reported in Table 13. The coefficient of variation (100 x $s/\bar{x}$) for these replicates varied from 8.6% for 20 ppb standards to 2.0% for 300 ppb standards. The 95 percent confidence interval for these analysis was 2 times the CV and varies from +17.2% for the 20 ppb standard and +4.0% for the 400 ppb standard. Accuracy for the TOX analyses was measured by the percentage of recovery of the known standards. This varied from 91% to 96% for these standards.

Multiple replicate analyses were also performed on several unknown samples. These data are given in Table 14. The coefficient of variation for these replicate varied from 4.8 to 8.2%.

Duplicate THM analyses are presented in Table 15. These values were used to determine analytical precision by calculation of I and the UCL for each duplicate pair. The average value for I was 0.047 and the standard deviation was 0.042. The upper confidence level and upper warning level were 0.17 and 0.13 respectively. This would indicate a 95% confidence interval of less than 20% for duplicate analyses. Of the 25 duplicates, none exceeded the UCL and one duplicate pair exceeded the UWL. Multiple replicate analyses for THM samples are presented in Table 16. The results of these analyses support the duplicate analysis results above. In all cases the 95% confidence interval was less than ±20 percent.

In addition to UCF, Lee County Environmental Laboratory also performed THM analyses. Data for duplicate analyses are presented in Table 17. None of the duplicate samples exceeded the UCL or UWL for I. The calculated distribution indicates that the 95% confidence interval for all four THM species is less than ±20%.

Table 11. Dissolved Organic Carbon Accuracy

	Analyst			
	1		2	
True Value (mg/L)	Found (mg/L)	% Recovery	Found (mg/L)	% Recovery
6.1	6.1	100	6.1	100
91.8	90.8	99	91.4	100

TABLE 12. THM Duplicates

A (ug/L)	B (ug/L)	I
94.2	96.4	0.0115
91.8	86.8	0.0280
87.4	86.2	0.0069
99.8	97.8	0.0101
87.4	93.6	0.0199
94.8	103.8	0.0453
91.8	96.6	0.0255
89.2	94.2	0.0273
101.4	94.2	0.0526
96.2	90.6	0.0300
90.0	95.6	0.0302
88.2	93.8	0.0308
89.4	88.4	0.0056
98.2	99.6	0.0071
89.4	92.0	0.0143
87.6	92.0	0.0245
90.0	99.0	0.0476
93.4	92.8	0.0032
192.4	177.8	0.0394
198.0	189.0	0.0233
294	284	0.0166
275	290	0.0272
292	289	0.0034
162	149	0.0418
210	188	0.0553
154	151	0.0098
166	182	0.0460
64	58	0.0492
207	178	0.0753 *
36.7	39.5	0.0367
45.9	48.6	0.0286
754	50.1	0.0475
1370	1482	0.0393
874	824	0.0294
1535	1552	0.0055

$\bar{x}$ = 0.0277
s = 0.0178
UCL = 0.0811
UWL = 0.0633
* = exceeds UWL

TABLE 13. Accuracy of TOX Analyses as Measured by Percent Recovery of a Known Standard

True Value (ug/L)	n	AV Found (ug/L)	s	Recovery	CV x 100
20.0	5	18.6	1.6	93	8.6
40.0	7	36.4	2.8	91	7.7
60.0	4	57.8	4.0	96	6.9
100.0	59	93.4	5.0	93	5.4
200.0	13	191	6.4	95	3.6
300.0	9	288	5.8	96	2.0
400.0	1	368	--	92	--

TABLE 14. Replicate TOX Analysis (ug/L as Cl)

n	$\bar{x}$	s	CV(100)
4	211	17.3	8.20
6	405	19.5	4.82
10	513	32.6	6.35

Table 15. THM's Precision

A (µg/L)	B (µg/L)	I
515	560	0.0564
195	140	0.1642*
155	175	0.0606
75	61	0.1029
69	58	0.0866
37	41	0.0513
138	138	0
120	141	0.0805
128	162	0.1172
1710	1760	0.0144
3480	329@	0.0281
3440	3290	0.0223
2180	2290	0.0246
2260	2480	0.0464
34	34	0
16	16	0
1314	12b2	0.0242
13	11	0.0833
11	11	0
18	17	0.0286
254	292	0.0696
23	23	0
181	209	0.0718
68	65	0.0226
265	228	0.0293

$\bar{x} = 0.047$
$s = 0.042$
$n = 25$
UCL = .1737
UWL = .1316
* = exceeds UWL

TABLE 16. THM Precision Data (ug/L)

n		$\overline{x}$	s	I 95% CI %
5	$CHCl_3$	44.2	1.79	±8%
	$CHCl_2Br$	54	1.58	±6%
	$CHClBr_2$	6.2	.57	±18%
5	$CHCl_3$	36.4	2.19	±12%
	$CHCl_2Br$	48.0	2.81	±12%
	$CHClBr_2$	5.1	.42	±16%

n	$\overline{x}$	s	95% CI %
3	271	17	±13%
3	328	29	±18%
4	299	11	±7%

TABLE 17. Lee County THM Analyses Duplicate Precision

$CHCL_3$			$CHBrCl_2$			$CHClBr_2$			$CHBr_3$		
A	B	I	A	B	I	A	B	I	A	B	I
8.4	7.9	.031	9.2	8.8	.022	8.6	8.2	.024	8.8	8.4	.0233
9.4	8.7	.039	9.8	10.8	.049	12.4	11.2	.051	12.6	11.8	.0328
15.3	13.9	.048	10.4	8.6	.095	10.6	9.8	.039	10.2	9.6	.0303
24.4	21.8	.056	12.6	10.2	.105	18.8	18.0	.022	18.4	17.1	.0366
18.6	19.1	.013	16.6	19.4	.078	24.2	26.6	.047	34.8	32.2	.0388
23.4	21.6	.040	17.2	17.8	.017	11.6	10.2	.064	26.4	24.2	.0435
28.6	30.7	.035	20.4	16.0	.121	28.4	29.6	.021	39.6	38.4	.0154
31.1	29.4	.028	20.8	22.6	.041	32.6	30.8	.028	14.2	16.2	.0658
19.6	21.2	.039	24.4	21.0	.075	24.6	21.8	.060	28.8	27.6	.0213
26.8	28.4	.029	26.2	29.6	.061	38.2	39.4	.015	36.4	35.8	.0083
38.7	41.4	.034	30.8	27.8	.051	14.8	13.4	.050	24.0	23.2	.0169
36.4	31.2	.077	32.0	31.2	.012	9.4	9.6	.011	3.4	8.8	.0330
26.4	29.1	.049	42.0	40.4	.022	16.6	15.2	.044	16.8	15.8	.0307
14.6	12.2	.089	39.8	43.4	.043	27.6	24.4	.062	187.4	17.8	.0166
18.8	17.3	.042	38.6	36.2	.032	33.2	36.0	.040	27.2	21.8	.0091
27.6	28.2	.011	18.2	22.2	.099	17.8	17.2	.017	30.4	29.8	.0010
	$\overline{x}$	.041		$\overline{x}$	.055		$\overline{x}$	.037		$\overline{x}$	.027
	s	.020		s	.034		s	.018		s	.016
	UCL	.101		UCL	.160		UCL	.091		UCl	.075
	USL	.081		USL	.126		USL	.073		USL	.059

Accuracy, as measured by percentage of recovery from spiked samples is shown in Table 18 for the Lee County Environmental Lab. The 95% confidence intervals for percentage of recovery were less than ±20% for 11 species except for dibromochloromethane ($CHClBr_2$) which had a 95% Cl of ±23%. In addition, Lee County has participated in 5 performance evaluations through the Florida Department of HRS. In all samples, percentage of recovery was within the range of 84-112%.

Several sets of samples were split between UCF and Lee County Environmental Lab for duplicate analyses (see Table 19). These samples were taken in the Lee County distribution system at the N. Reservoir and Eland sites. Overall, the I statistic for these duplicates was 0.055 and the standard deviation for I was 0.040. The UCL was 0.175 and UWL was 0.135. The I statistic for all 59 duplicate pairs was less than UCL and for 3 of the duplicate pairs I exceeded the UWL.

Data for the precision of Lee County's aluminum analyses is presented in Table 20. The average and standard deviation of I were 0.030 and 0.033 respectively. The upper warning level was 0.096 and the upper control level was 0.129. One duplicate pair exceeded the upper control level. The sample had a concentration of less than 0.1-mg/L and it is likely that the distribution of I is different for samples in this range. Aluminum accuracy is shown in Table 21. Percentage of recovery of spiked samples varied from 92 to 103%. The 95% confidence interval for percent recovery was 91-105%.

Aluminum residuals were measured at UCF using a Beckman SpectraSpan V inductively coupled plasma emission spectrometer. Precision and accuracy for this instrument were determined by analyzing replicate solutions of known concentration. These data are presented in Table 22. Percentage of recovery varied from 95 to 104.

5.5.2 EPA Performance Evaluation

Two separate project evaluations were performed for two samples each. The results of those evaluations are shown in Table 23. The results from the first evaluation indicated that three of the THM samples were found to be outside of an acceptable range. This problem was traced to lack of low level sensitivity relative to the levels of THM normally encountered at the project sites, 300-1800 μg/L.

Additional control samples were requested and received from EPA. Subsequent analyses indicated the typical G-C settings for the higher levels of THM's were not sensitive enough for the EPA samples. This problem was corrected by increasing purge gas flow and decreasing attenuation for EPA performance evaluation samples. The results of the 1/04/85 EPA performances samples were satisfactory after these changes were incorporated.

Lee County Environmental Laboratory participated in THM analysis performance evaluations sponsored by the Florida Department of HRS. Results of these analyses are given in Table 24. Analyses were within

TABLE 18. Lee County Analyses Percent Recovery

$CHCl_3$			$CHBrCl_2$			$CHClBr_2$			$CHBr_3$		
Known	Found	%R	Known	Found	%R	Known	Found	%R	Known	Found	%R
8.4	7.9	84	9.2	9.8	107	10.6	8.6	81	8.4	9.0	107
8.4	8.9	106	9.2	10.0	109	10.6	9.0	85	8.4	8.8	105
9.2	8.6	94	8.8	8.2	93	14.8	14.2	96	12.2	13.2	108
9.2	8.8	96	8.8	8.6	98	14.8	16.4	111	12.2	12.6	103
12.4	13.0	105	12.4	12.8	103	9.2	10.8	117	20.6	19.2	93
12.4	12.8	103	12.4	11.8	95	9.2	11.0	120	20.6	18.4	89
16.2	14.8	91	18.6	19.2	97	16.4	19.2	117	36.4	38.2	105
16.2	15.3	94	18.6	19.8	107	16.4	18.6	113	36.4	35.8	98
21.6	18.9	88	24.4	23.4	96	24.2	23.6	98	30.8	28.6	93
21.6	19.6	98	24.4	22.6	93	24.2	24.0	99	30.8	28.6	92
26.8	23.4	87	32.2	34.4	107	32.6	29.4	90	26.6	23.8	90
26.8	24.6	92	32.2	33.0	103	32.6	29.4	90	26.6	24.2	91
31.2	27.2	87	38.8	41.6	107	38.2	36.4	95	14.6	14.2	97
31.2	28.4	91	38.8	39.4	102	38.2	39.6	104	14.5	15.4	106
34.6	38.6	90	42.8	40.6	95	40.0	40.8	102	10.2	9.2	90
34.6	32.4	94	42.8	41.4	97	40.0	38.2	96	10.2	8.8	86
38.8	34.4	89	46.4	47.2	102	24.6	22.2	90	22.4	20.2	90
38.8	42.2	109	46.4	44.6	96	24.6	23.4	95	22.4	21.4	96
	$\overline{x} = 94$			$\overline{x} = 100$			$\overline{x} = 100$			$\overline{x} = 97$	
	$s = 6.7$			$x - 5.1$			$s = 11.6$			$s = 7.2$	
95% CI =	81-108			90-110			77-123			82-111	

TABLE 19. THM Precision - Duplicate Samples Analyzed at UCF and Lee County

Lee County (ug/L)	UCF (ug/L)	I (ug/L)
296	292	0.0068
243	281	0.0725
254	295	0.0747
357	395	0.0505
321	313	0.0126
197	203	0.0150
212	195	0.0418
187	207	0.0508
190	216	0.0064
190	210	0.0050
202	243	0.0921
228	247	0.0400
217	261	0.0921
239	290	0.0964
211	251	0.0866
240	286	0.0875
235	284	0.0944
242	237	0.0104
232	243	0.0232
259	277	0.0366
255	256	0.0039
261	190	0.1574
257	244	0.0259
257	284	0.0499
201	206	0.0123
231	231	0
252	242	0.0202
339	389	0.0687
230	306	0.1418*
391	423	0.0393
345	354	0.0129
236	237	0.0021
206	194	0.0300
217	198	0.0458
175	218	0.1094
202	246	0.0982
218	268	0.1029
204	234	0.0685
236	278	0.0817
268	254	0.0268
243	265	0.0433
232	284	0.1008
240	256	0.0323
247	324	0.1349
280	344	0.1026
266	305	0.0683
241	263	0.0437
258	288	0.0409
258	288	0.0549
294	273	0.0370
260	274	0.0262
309	277	0.0546
392	263	0.0523
257	287	0.0551
257	285	0.0517
217	298	0.1573
255	261	0.0116
238	248	0.0206

n = 59
$\bar{x}$ = 0.0551
s = 0.0399
UCL = 0.1747
UWL = 0.1348
* = exceeds UWL

TABLE 20. Aluminum Precision Lee County

A (mg/L)	B (mg/L)	I
0.06	0.09	0.200 *
0.45	0.46	0.011
0.80	0.87	0.042
1.24	1.14	0.042
1.76	1.91	0.041
2.93	3.11	0.039
3.68	3.46	0.031
4.21	4.39	0.021
5.33	5.28	0.005
5.92	5.99	0.006
6.11	6.31	0.016
6.54	6.36	0.014
7.22	7.09	0.009
8.44	8.56	0.007
8.16	8.39	0.014
9.47	9.39	0.053
9.88	9.76	0.006
10.10	9.94	0.008
10.41	10.63	0.010
10.63	10.31	0.015

$\overline{x}$ = 0.0300
s = 0.0329
UCL = 0.1287
UWL = 0.0958
* = exceeds UCL

TABLE 21. Lee County Aluminum Accuracy

Known (mg/L)	Found (mg/L)	%R
1.00	0.93	93
1.00	0.97	97
1.00	1.00	100
1.00	0.92	92
3.00	2.96	99
3.00	3.15	105
3.00	3.00	100
3.00	2.87	96
5.00	4.81	96
5.00	5.06	101
5.00	4.90	98
5.00	4.71	94
8.00	7.81	98
8.00	8.14	102
8.00	8.21	103
8.00	7.87	98
10.00	9.24	92
10.00	9.67	97
10.00	10.10	101
10.00	9.92	99

$\bar{x} = 98.1$
$s = 3.56$

TABLE 22. Aluminum Recovery for ICP Analysis

Standard (mg/L)	1.00		2.50		5.00	
	mg/L	% Recovery	mg/L	% Recovery	mg/L	% Recovery
Found Values (mg/L)	1.04	104	2.43	97.2	5.04	100.8
	1.01	101	2.41	96.4	5.09	101.8
	1.03	103	2.41	96.4	4.97	99.4
	1.00	100	2.39	95.6	4.99	99.8
	1.02	102	2.38	95.2	5.00	100.0
AV mg/L	1.00	102.0	2.40	96.2	5.02	100.4
S mg/L	0.02	1.6	0.02	0.8	0.05	1.0

Table 23. EPA Performance Evaluation Summary

	Date 7/16/84		1/04/85	
Sample Parameter	1	2	1	2
$CHCl_3$-µg/L				
Reported	13.6	58.7*	86.6	34.4
True	11.6	47.5	89.0	35.6
range	9.52 - 14.3	38.0 - 57.0	71.2 - .07	28.5 - 42.7
$CHBr_3$-µg/L				
Reported	20.5	53.0	31.7	13.8
True	20.3	50.8	33.8	13.5
range	16.2 - 24.4	40.6 - 01.0	27.0 - 40.6	10.8 - 16.2
$CHCl_2BR$-µg/L				
Reported	11.5*	89.4*	45.5	14.8
True	9.40	70.5	47.0	14.1
range	7.52 - 11.3	56.4 - 84.6	37.6 - 56.4	11.3 - 16.9
$CHBR_2Cl$-µg/L				
Reported	20.0	63.5	73.5	25.0
True	16.7	61.3	72.5	22.3
range	13.4 - 20.0	49.0 - 73.6	58.0 - 27.0	17.8 - 26.8
Free Cl_2-mg/L				
Reported	1.00	0.15*	0.65	0.90
True	1.51	0.433	0.800	1.10
range	0.972 - 2.00	0.201 - .630	7.12 - 9.29	.683 - 1.46
Turbidity - NTU				
Reported	0.41	1.90	7.65	0.98
True	0.50	2.0	8.0	0.8
range	0.365 - 879	1.73 - 2.59	7.12-9.29	0.645 - 1.18
Total THM-µg/L				
Reported	65.6	264.6	237.3	88.00
True	58.3	230.1	242.3	194 - 291
range	46.6 - 70.0	184 - 276	194 - 291	68.4 - 103

* Outside of range.

TABLE 24. Lee County Performance Evaluation

	WS 009		WS 011		WS 013		WS 014		WS 015	
	1	2	1	2	1	2	1	2	1	2
$CHCl_3$ - ug/L										
Reported	28.5	105	85.3	12.8	72.3	22.8	11.1	46.8	87.3	32.4
True	25.5	102	86.8	15.3	71.2	23.7	11.9	47.5	89.0	35.6
% Recovery	112	103	98	84	102	96	93	99	98	91
$CHBr_3$ - ug/L										
Reported	12.2	67.1	56.2	16.8	54.8	27.8	21.1	51.5	35.9	13.4
True	11.0	65.8	54.9	16.5	50.3	27.1	20.3	50.8	33.8	13.5
% Recovery	111	102	102	102	109	103	104	101	106	99
$CHCl_2Br$ - ug/L										
Reported	14.3	57.1	84.7	8.8	59.0	17.1	9.6	69.8	47.2	13.0
True	13.7	59.3	82.1	9.12	56.4	18.8	9.4	70.5	47.0	14.1
% Recovery	104	96	103	96	105	91	102	99	100	93
$CHBr_2Cl$ - ug/L										
Reported	5.8	81.4	108	17.8	40.2	10.2	17.0	62.2	73.5	20.9
True	5.92	88.9	107	17.8	39.0	11.2	16.7	61.3	72.5	22.3
% Recovery	98	92	101	100	103	91	102	101	101	94

acceptable range (±20% of the true value) for all parameters. These results confirm previous precision and accuracy tables indicating that THM analyses were within accpetable control levels.

5.6 Summary

All analytical methods used in the project followed Standard Methods except those described in this chapter. These included TOX, TOXFP, THM's, and color. Precision and accuracy for TOX, THM, aluminum and DOC are presented. Precision is measured using the industrial statistic, I, as prescribed in the EPA Manual for Analytical Quality Control in Water and Wastewater Laboratories. This statistic is the ratio of the standard deviation for duplicate analyses to their mean. Control limits for DOC, TOX, THM, and aluminum were ±13%, ±8%, ±17%, and ±13% respectively. Control limits for samples analyzed at the Lee County Environmental Lab were ±16% and ±13% for THM and aluminum respectively. Accuracy, as measured by recovery of spiked samples, was evaluated in EPA performance evaluations. Initial procedural problems with low THM values analyses were experienced. These problems were corrected and subsequent performance was within control limits. Spiked TOX samples indicated the TOX recovery was within ±17%. Aluminum residuals were analyzed with an accuracy of ±8%.

The overall quality of analytical results for this project was within acceptable limits. Analytical problems that did occur were detected by the quality measurements and corrected.

CHAPTER 6

RESULTS AND DISCUSSION

6.0 Introduction

This chapter presents and discusses the results from the bench, pilot, and plant scale investigations of softening, coagulation, membranes, and flotation for THM control. It is organized to present the results of the plant optimizations in order to present a base line for the new technology investigations at each site. The PACl investigations at Olga are then presented followed by the sequential precipitation work at Olga and AID (a surface and ground water source). This order provides a logical presentation of data that allows a comparison of all precipitative processes (coagulation and softening).

Membrane processes are presented next. First the small scale pilot plant work at VOG, AID, and Olga is presented followed by the large scale pilot plant at VOG and AID. This order allows membrane processes to be related to a surface (Olga) and ground water (VOG/AID) source. Also membrane treatment capability can be compared to precipitative processes.

Finally the bench scale flotation results are presented. The flotation work was never extended to a pilot plant scale because of the poor THM precursor removal in the bench operations.

6.1 Plant Optimization

After the four sites for demonstration of new technology were selected, each of the plants was optimized for THM precursor removal. This involved first a laboratory study to determine the capability of the plant to remove organics. Second, a presentation was made to site management to convey the results and to obtain their permission to optimize their process. Then the plant scale tests were conducted at AID, VOG, Venice, and Olga. The Olga results are presented in the PACl section because that plant optimization involved alum coagulation whereas AID, VOG, and Venice had lime-softening operations. The intent of plant optimization was to determine the degree and cost of optimizing the existing plant process for THM precursor removal.

6.1.1 AID Plant Optimization

6.1.1.1 Introduction

ACME Improvement District Utilities is located in Palm Beach County, Florida approximately 10 miles west of West Palm Beach. The utility serves a predominantly residential service area with some commercial usage. Currently the plant serves approximately 4000 connections with an estimated population of 10,000.

The raw water supply is obtained from a series of wells which tap a local, shallow, unconfined aquifer. Recharge of this aquifer is through rainfall and surface water (predominantly canals) infiltration. The representative water quality is given in Table 25.

The plant utilizes the lime-softening process. The treatment scheme is presented in Figure 4. Basically, the raw water is aerated and then softened and settled in a solids contact unit by adding lime (230-mg/L CaO) and a coagulant polymer (Olin 5011). The settled water is then chlorinated (20 mg/L Cl_2), filtered and stored in above-ground storage tanks. On certain occasions, a small chlorine dose (1.0-mg/L) is applied in the storage tanks to allow a residual. The finished water quality is shown in Table 25.

6.1.1.2 AID Jar Test

The purpose of this phase was to optimize the removal of organics by adjusting the pH of the lime-softening process on a bench scale. Such pH adjustments were made by lime. In the formation and settling of $CaCO_3$ and $Mg(OH)_2$ flocs from low turbidity waters, organic compounds such as humic and fulvic acids are removed via "sweep coagulation." By increasing the pH, the volume of floc formed is generally increased and possibly a greater organics removal should occur due to the increased opportunity for adsorption or the precipitation of organo-metallic components at higher pH.

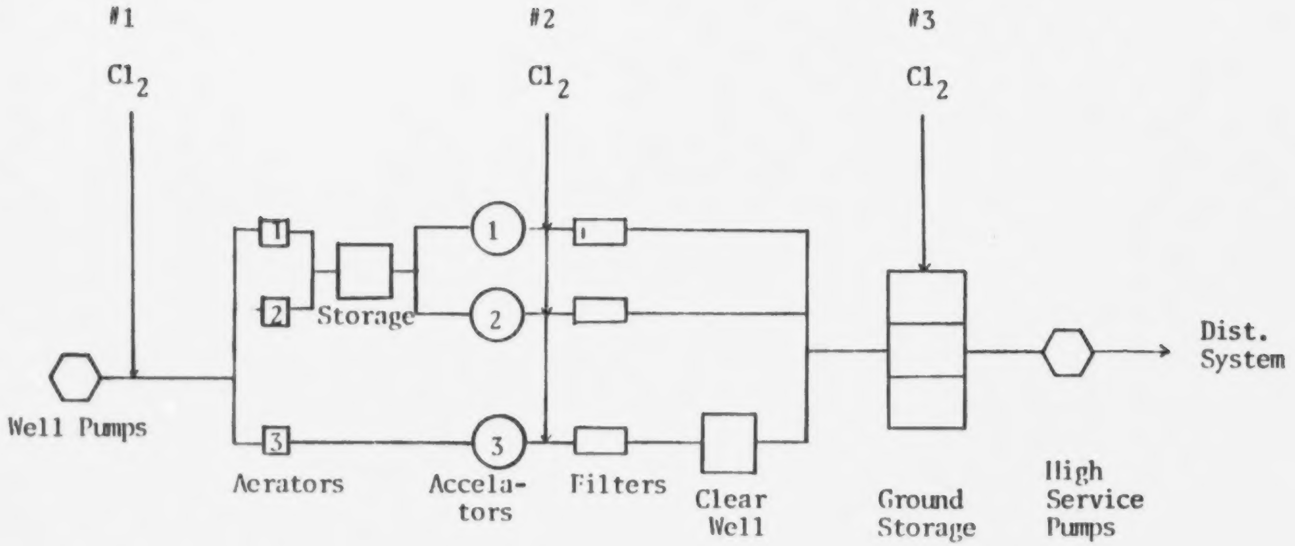
The tests were conducted using one-liter jar tests on raw waters collected from the aeration unit. The water was stored for less than two weeks at 4°C to prevent biological and chemical alteration. The order of chemical addition was lime, followed by polymer with rapid and slow mix times of one and 20 minutes, respectively. The results of the test may be seen in Figure 5. The exact data is recorded in Table A-1.

As can be seen from the data, the organic parameters tend to drop in concentration with increasing pH. Color and DOC reach minimums at pH 11.5 while THM's was minimized at pH 11.0. Based on these results, pH 11.3 was chosen for further testing. This pH was chosen since good organic parameter removal was achieved. In addition, this pH could be obtained without excessive chemical doses. Most plants soften at pH 10.3. This data suggests that 5% to 10% more DOC and more than 30% additional color could be removed at pH 11.0. However, this would correspond to an increased hardness of 40-mg/L $CaCO_3$ from the softening reactor. There was a general trend to decrease color, DOC, and THM's linearly from pH 7.5 to 11.5. This

Table 25. Acme Improvement District Water Quality

Parameter	Raw	Finished
pH	7.20	8.75
DOC (mg/L as C)	12.20	9.00
TOX (µg/L as Cl)	20	750
THM (µg/L)	ND	275
Color (cpu)	35	6
TDS (mg/L)	470	240
Total Hardness (mg/L as $CaCO_3$)	340	100
Calcium Hardness (mg/L as $CaCO_3$)	310	70
Conductivity (mmhos)	0.620	0.320

ND = Not Detectable


#1
Cl_2
#2
Cl_2
#3
Cl_2
1
2
3
1
2
3
Storage
Well Pumps
Aerators
Accela-
tors
Filters
Clear
Well
Ground
Storage
High
Service
Pumps
Dist.
System

Cl_2 - #1 Not in use
Cl_2 - #2 20 mg/1
Cl_2 - #3 1 mg/1

Dosages in Accelators
Lime = 222 mg/1
Olin 5011 = 0.15 mg/1

Unit #1 = 0.65 MGD
Unit #2 = 1.00 MGD
Unit #3 = 2.00 MGD

Figure 4 Flow Schematic for AID Water Treatment Plant

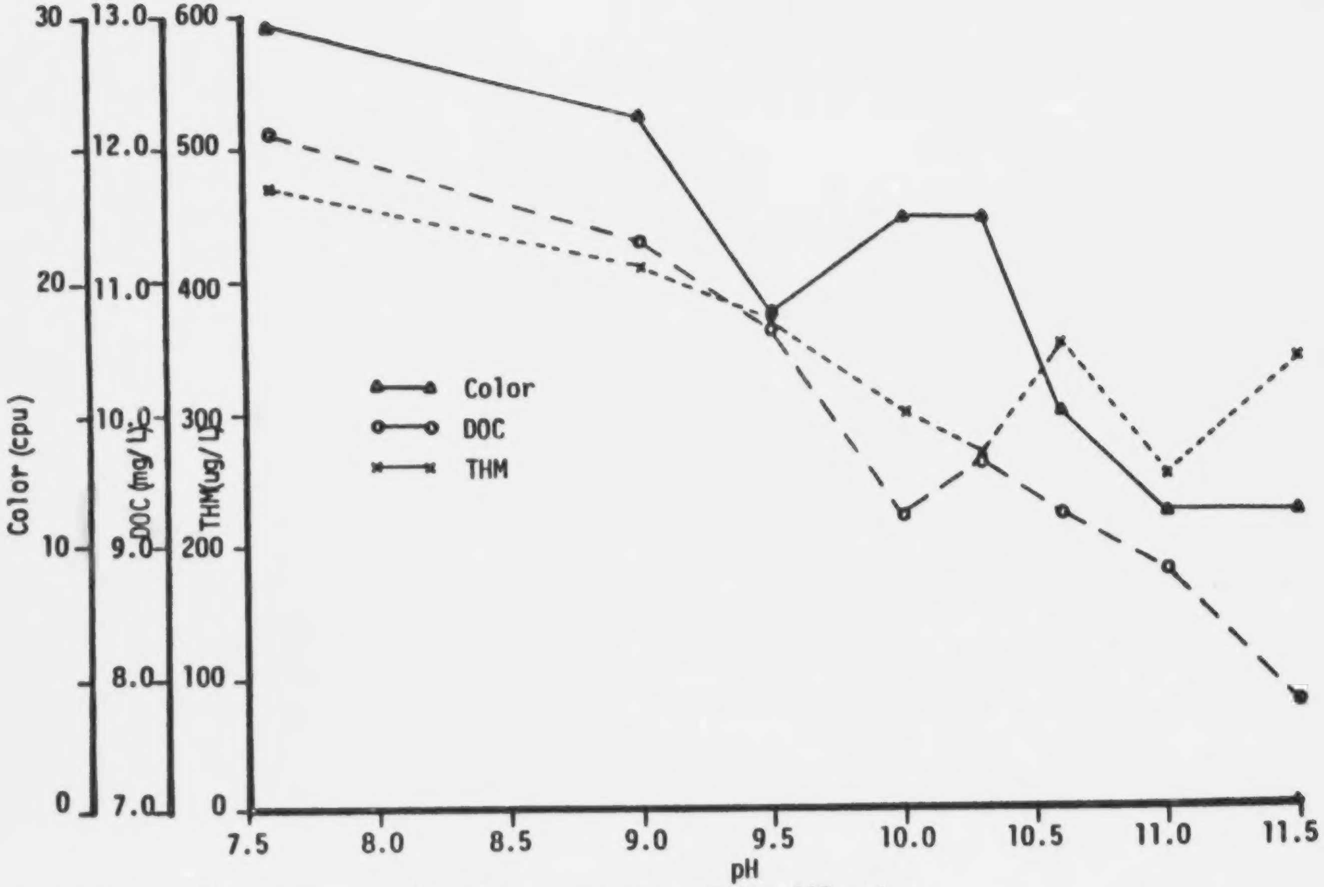


Figure 5 DOC, Color and THMFP vs. softening pH for AID water.

decrease approximately corresponded to 66% of the initial color, 36% of the initial DOC, and 25% of the THM's. Over the softening pH ranges of 9.0 to 11.5, a general statement could be made that a 1% to 2% increase for the removal of the color, DOC, THM's, and TOXFP could be expected for every 0.1 pH unit increase in the softening reactor if all other parameters (Cl_2, temperature, etc.) are held constant.

6.1.1.3 Coagulant Aid

In this phase of the study, three coagulants were examined in conjunction with lime-softening at pH 11.3 with the intent of increasing organic removal. The coagulants were commercial alum and two polyaluminum chloride solutions, PACl 190 and PACl 100S.

The tests were conducted using one liter jar tests with coagulant added first so that it might react with the organics at a lower pH prior to lime softening. The coagulant doses were equal to 0.8, 1.6, 2.4 and 4.0-mg/L Al. The data for DOC, color, THMFP, and Al residuals may be seen in Figures 6, 7, 8, and 9. A control jar value was obtained by lime softening at pH 11.3 without any coagulants.

From the data, several observations may be made:

1. All the coagulant aids generally reduced DOC as compared to the control with alum giving the best overall reduction. This removal appears to be realized at the minimum Al dose.
2. No discernible trends were noted for color removal. However, all coagulants at all doses gave a finished water color of less than 10 cpu.
3. All the coagulants at all doses caused a minor decrease in THM's. No one coagulant appeared significantly superior to any other.
4. Alum clearly gave the lowest aluminum residual at all applied dosages.

From the above noted trends, alum at an equivalent aluminum dose of 0.8-mg/L was chosen for further testing. The decision for this coagulant and dose were as follows:

1. This coagulant and dose was the optimum for DOC removal and second best for THM's.
2. The residual color from this coagulant and dose was below 10 cpu.
3. This dose exhibited a low aluminum residual which would be advanageous if MCLs are set for aluminum in the future.
4. Alum enjoys a cost and familiarity of use advantage over PACl. In addition, the availability in this country is greater.

The data indicates that use of alum as a coagulant aid at AID could increase organic removal by 8% to 15%. This would correspond to an expected THM decrease of 10% to 20% but may affect floc settling and sludge filterability.

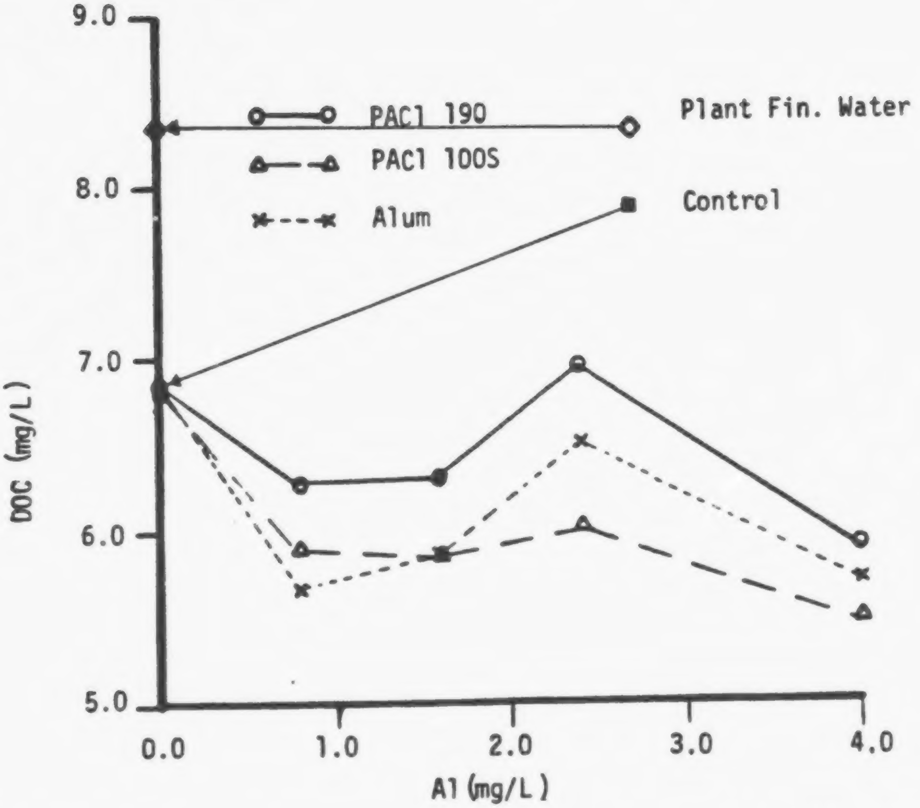


Figure 6 DOC remaining after lime softening in combination with Al polymer for AID water.

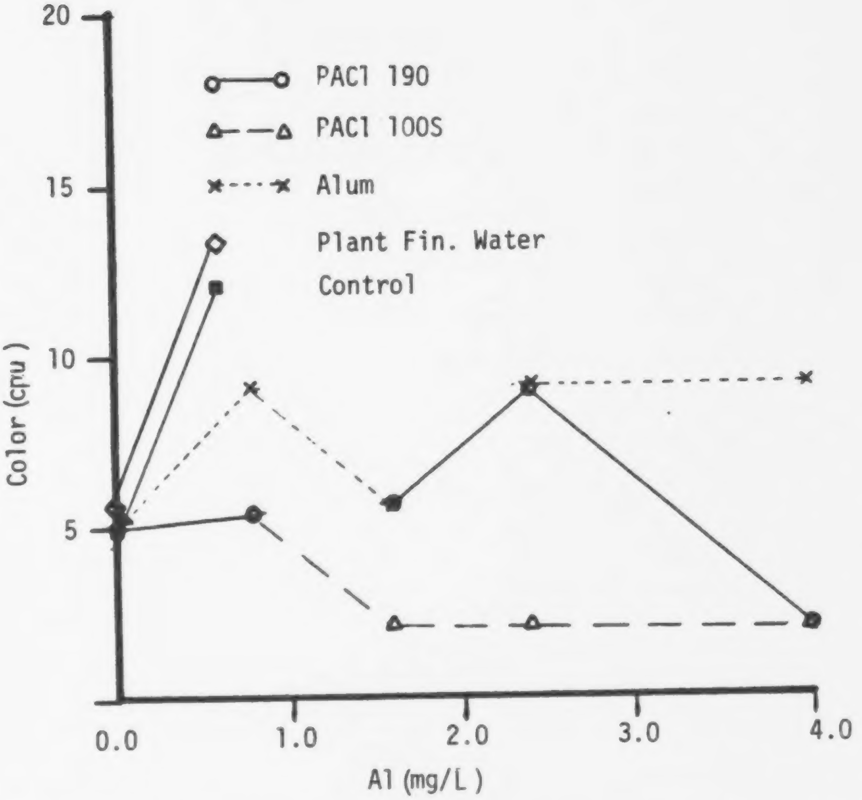


Figure 7 Color remaining after lime softening in combination with Al polymer for AID water.

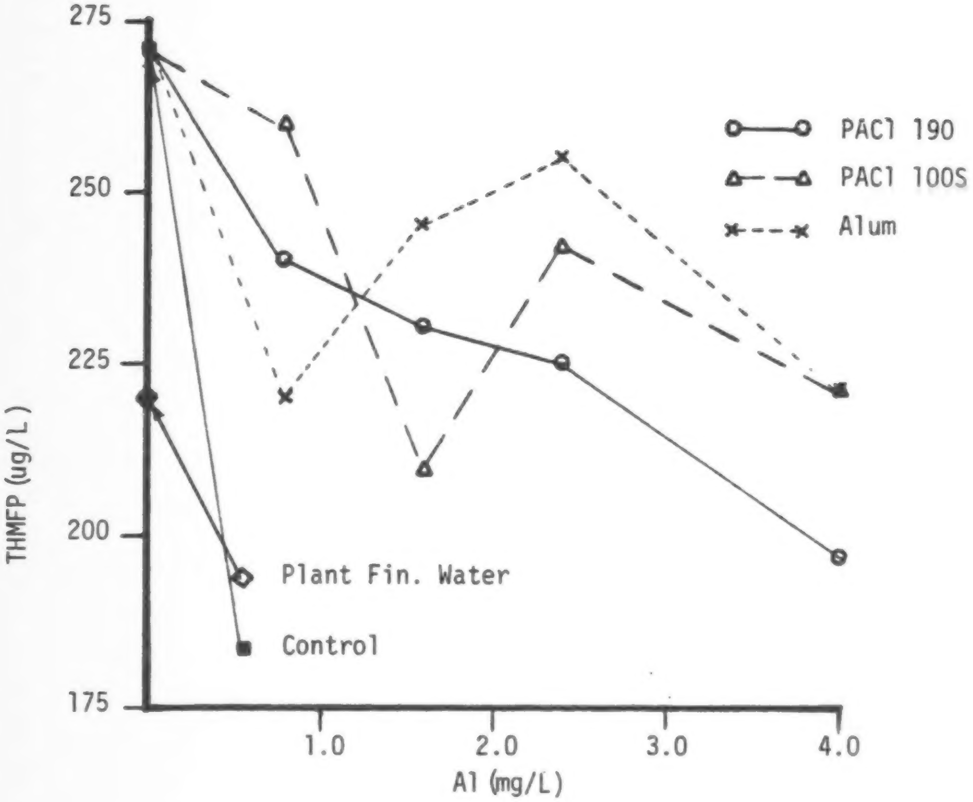


Figure 8 THMFP remaining after lime softening in combination with Al polymer for AID water.

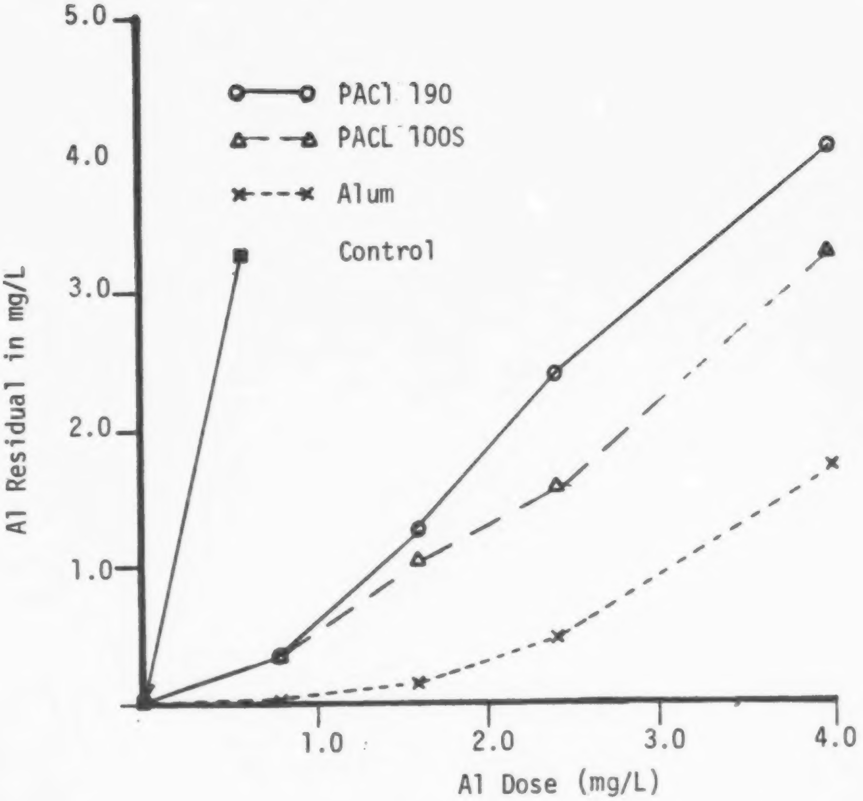


Figure 9 Al residual after lime softening in combination with Al polymer for AID water.

6.1.1.4 Chlorine Dissipation and THM Formation

Once the experimental coagulant dose and softening pH were established, it was necessary to compare the UCF product water against a finished water using plant chemical doses. A series of experiments were conducted to determine required Cl_2 dose for a 96 hr. residual and color removal to less than 15 cpu following softening.

In order to compare residual color after 30 minutes contact times, both UCF and plant water were treated with various chlorine doses. The results are given in Figure 10. As can be seen, a chlorine dose of 2.5-mg/L was required to bring the plant water to the 15 cpu color limit, while the UCF water was below 15 cpu without the addition of chlorine. The rate of color removal at colors >5 cpu is notably higher relative to color <5 cpu.

The Cl_2 dissipation study served two different purposes. The first was to determine the chlorine dose for the UCF water that would still allow a residual after 96 hours. Second, to directly compare the plant water chlorine dissipation rate with the UCF rate.

The results from these tests may be seen in Figure 11. As can be seen, the 10-mg/L Cl_2 dose will still allow a free chlorine residual of 0.20-mg/L for almost 96 hours. Since the detention time in the system is probably three days or less, a chlorine dose of 10-mg/L was chosen for further tests. It should be further noted that for identical initial doses (20-mg/L) the UCF water had a substantial greater chlorine residual throughout the 96 hour test.

To compare the THM formation potentials of the two waters, a 96 hour THM formation curve was established. It was important to note that an incubation temperature of 30°C was used. This temperature is approximately 10°C warmer than the temperatures found in the system. The 30°C temperature is not uncommon in Florida distribution systems and gives a "worst-case" THM value. Temperatures of finished waters in Florida range from 20°C to 30°C. A conservative approach to generating THM's in a laboratory utilizes conditions encountered in distribution systems that produce maximum THM's. The results for these tests can be seen in Figure 12. The UCF water produced a THM's value that was only one-half the plant THM's value.

6.1.1.5 Recommendations

Based on the above results, the following recommendations were approved by AID management for plant scale tests.

1. A minimum five-day continuous plant scale test conducted at the dosages established in the bench scale tests.
2. For pH control, temporary H_2SO_4 feed line prior to the chlorination point.

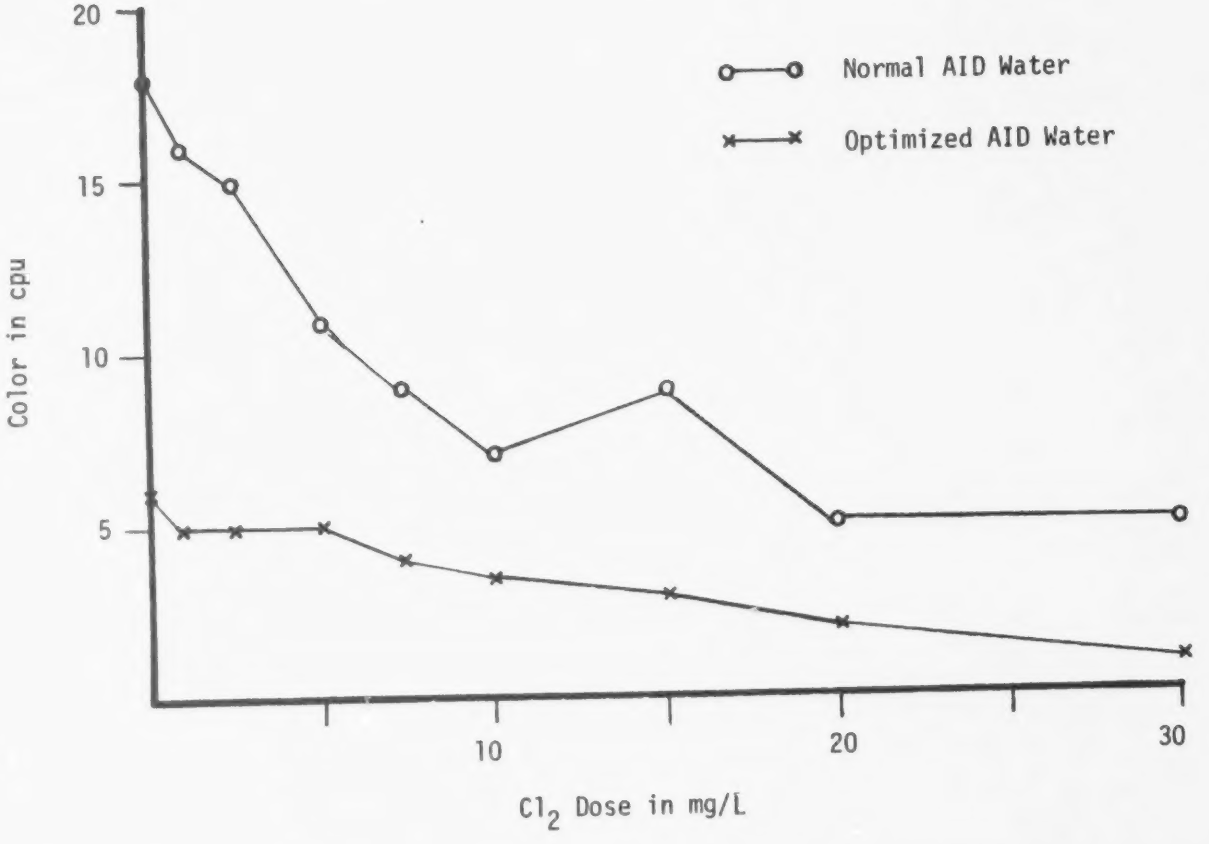


Figure 10 ACME Improvement District - Color vs. Cl_2 Dose After 30 Minute Contact Time.

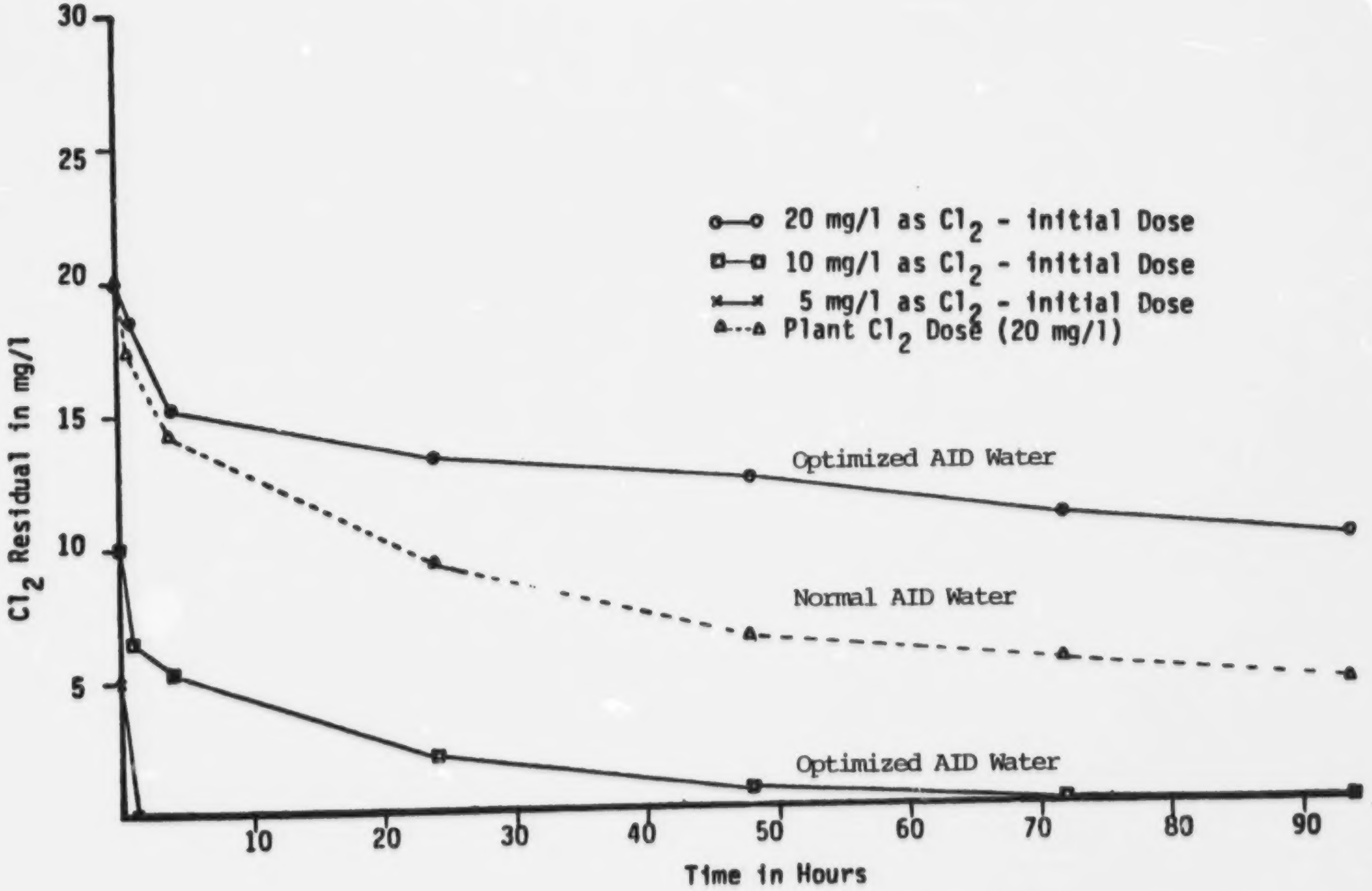


Figure 11 ACME Improvement District - Chlorine Dissipation Curve at pH 8.8 and 30°C.

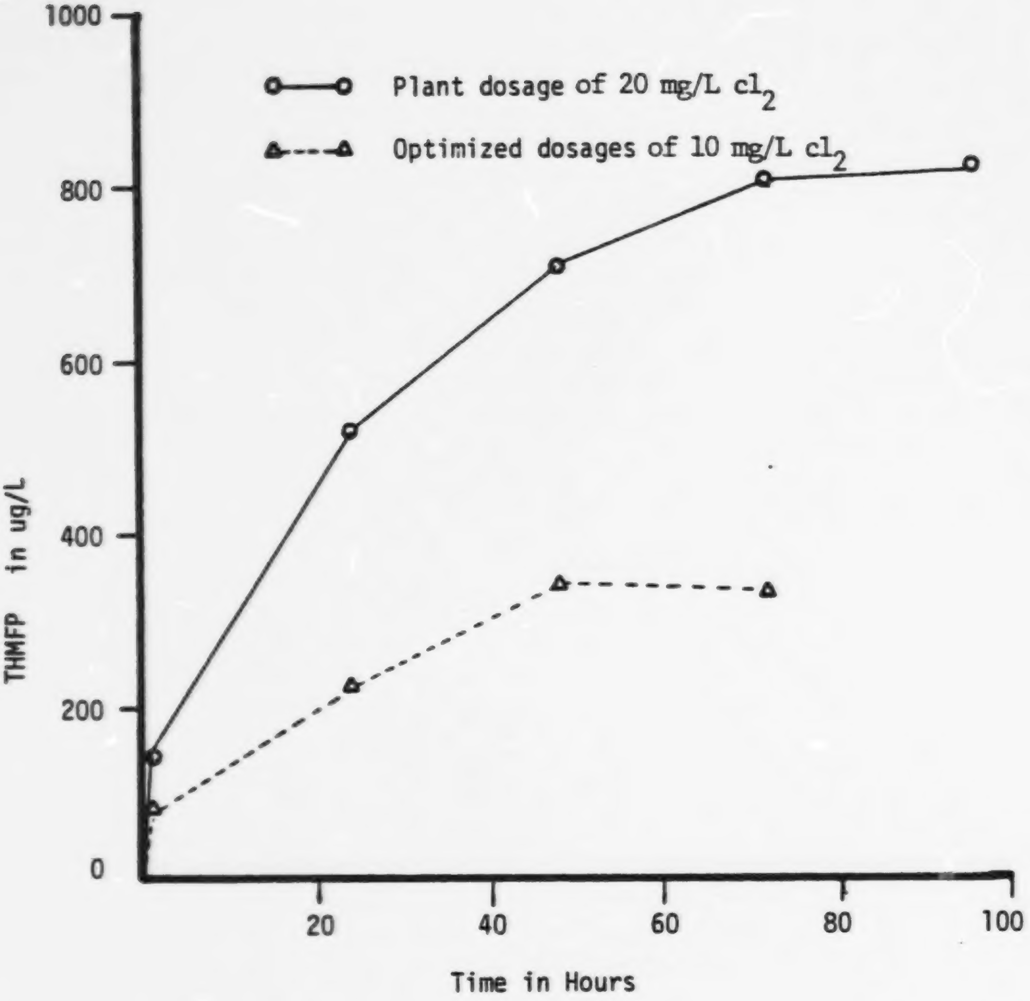


Figure 12 ACME Improvement District. THM Formation Curve for normal and optimized waters at pH 8.8 and 30°C.

3. Water samples during the test period collected daily at the following points: the raw water, the settled water prior to chlorination, the filtered water subsequent to chlorination, and a point within the distribution system. The samples analyzed for DOC, color, THM, aluminum residual, conductivity, pH, and other appropriate parameters.

6.1.1.6 AID Plant Optimization Project

Subsequent to the bench-scale studies which were completed in June, a plant-scale demonstration was initiated on August 1 and terminated on August 8. The project initially involved the increasing of the lime dose to obtain a softening pH of 11.3, the addition of 1.0 mg/L alum and a lowering of the chlorine feed to 10 mg/L. On the third day the alum feed was discontinued due to persistent turbidity problems in the softener. Also commencing on the third day was a rise in the chlorine feed to 14 mg/L in the pre-chlorination point and a post-chlorination feed of 3 mg/L. The increases were necessitated by the lack of residual in the storage tanks and distribution system under previous doses of 10, 12 and 14 mg/L but still represented a 3 to 8-mg/L decrease from normal plant operation. Plant operation was closely supervised by project personnel for the entire study period. However on August 8, the last day of operation, all chemical feeds were made by regular plant personnel.

The AID plant is rated at 3.5 MGD and produced 1.4 MGD, but the average hours of operation per day was 20.1. The plant was designed using conventional surface loading rates (SLR) of 0.5 gpm/ft^2 and 2.0 gpm/ft^2 for settling and filtration. Because the plant was operated lower than capacity, the design SLRs were not exceeded and no hydraulic stress was placed on the filters. The average detention time in the 2.0 million gallon reservoir was about 1.3 days.

The water quality through the AID plant is presented in Table 26. The DOC in the raw water varied from 15.58-mg/L on 8/5 to 12.84-mg/L on 8/1. A raw color variation of 42 to 34 cpu is also noted in Table 27. This variation in organic quality is somewhat surprising for a groundwater source, but it should be remembered that AID uses several wells that are less than 100' deep. Such shallow wells do have a variable water quality among and within each well.

The softening pH, denoted as launder, was increased from 10.3 on the first day to 11.2 however was generally held near 11.0 during the plant test. The increased pH from 10.3 to 11.0 was accompanied by a decreased alkalinity in the storage tank and distribution system from approximately 30 to 10-mg/L $CaCO_3$. This low alkalinity adds very little buffering to the water and could present a corrosion problem. Also from the high CaO dose to increase the reaction pH, an increase in final hardness is noted. The total hardness increased from 76-mg/L $CaCO_3$ at the beginning of the plant study to about 150-mg/L $CaCO_3$ at the end. This low alkalinity, high hardness finished water is typical of any lime softened water at pH 11 or more. Additional alkalinity (25-mg/L) should be added after chlorination in order to buffer the final water and maintain the good organic removal at pH 11.

Table 26. AID Plant Optimization Water Quality Data

Parameter	Sample Location	8/1	8/2	8/3	8/4	8/5	8/6	8/7	8/8	8/9	8/10
DOC (in mg/L as C)	Raw	12.84	13.08	13.14	14.83	15.58	14.47	14.64	13.67		
	Launder	7.54	6.41	6.29	6.20	7.65	7.26	7.27	6.66		
	Filter Top	8.44	7.12	8.03	7.09	7.66	7.85	7.89	7.65		
	Filter Bottom	--	--	--	--	--	--	--	--		
	Clear Well	6.89	8.23	7.15	7.38	7.05	7.73	6.55	6.19		
	Storage Tank	9.46	8.26	7.61	7.94	7.50	8.02	7.78	7.51	9.30	8.54
	Distribution System	8.97	8.72	7.81	8.78	7.42	8.09	7.52	7.75		
Color (in cpu)	Raw	34	39	42	35	40	38	40	39		
	Launder										
	Filter Top										
	Filter Bottom										
	Clear Well										
	Storage Tank	7.0	6.0	4.0	5.0	2.0	2.0	6.0	3.0	4	4
	Distribution System	11	8.0	4.5	3.0	1.0	2.0	4.0	3.0		
THM (in μg/L)	Raw	0	0	0	0	0	0	0	0		
	Launder	390	279	174	100	124	120	126	47		
	Filter Top										
	Filter Bottom										
	Clear Well										
	Storage Tank	440	248	286	191	270	238	188	227	251	365
	Distribution System	387	336	319	238	279	179	175	214		
TOX (in μg/L as Cl)	Raw	0	0	0	0	0	0	0	0	-	-
	Launder	-	-	-	-	-	-	-	-	-	-
	Filter Top	-	-	-	-	-	-	-	-	-	-
	Filter Bottom	-	-	-	-	-	-	-	-	-	-
	Clear Well	-	-	-	-	-	-	-	-	-	-
	Storage Tank	660	418	392	399	493	485	519	581	-	-
	Distribution System									394	489

(continued)

Table 26 (continued). AID Plant Optimization Water Quality Data

Parameter	Sample Location	8/1	8/2	8/3	8/4	8/5	8/6	8/7	8/8
Calcium Hardness (mg/L as $CaCO_3$)	Raw	252	192	336	304	300	308	304	
	Launder	64	148	138	168	136	140	156	
	Filter Top	64	160	142	174	140	140	160	
	Filter Bottom	64	152	146	174	136	140	160	
	Clear Well	60	146	120	176	152	136	160	
	Storage Tank	64	124	138	144	144	144	144	
	Distribution System	60	80	118	136	140	144	140	
Total Hardness (mg/L as $CaCO_3$)	Raw	260	212	340	312	312	328	320	
	Launder	76	158	138	170	144	144	166	
	Filter Top	76	164	144	178	148	148	164	
	Filter Bottom	76	166	148	178	144	148	164	
	Clear Well	68	152	126	180	160	144	164	
	Storage Tank	76	134	140	146	156	148	160	
	Distribution System	76	100	130	144	144	148	144	
Chlorine Residual (mg/L as Cl_2)	Raw	0	0	0	0	0	0	0	
	Launder	-	-	-	-	-	-	-	
	Filter Top	4.5	>3.5	3.4	>3.5	2.4	>3.5	>3.5	
	Filter Bottom	4.5	>3.5	3.0	>3.5	1.5	2.3	3.5	
	Clear Well	4.8	>3.5	2.7	3.5	1.4	1.5	2.0	
	Storage Tank	2.0	0.7	1.9	1.9	1.8	2.2	1.2	
	Distribution System	0.6	0.5	0.4	0.0	0.4	0.5	0.4	

(continued)

Table 26 (continued). AID Plant Optimization Test - Inorganic Parameter Values

Parameter	Sample Location	8/1	8/2	8/3	8/4	8/5	8/6	8/7	8/8
pH	Raw	7.4	7.0	6.9	6.9	6.9	7.1	7.0	
	Launder	10.3	11.2	11.0	11.1	10.9	10.9	10.9	
	Filter Top	9.2	10.1	10.5	10.0	8.8	9.2	10.1	
	Filter Bottom	8.9	10.3	10.5	10.1	8.5	8.0	9.9	
	Clear Well	8.8	10.3	10.1	10.2	8.6	7.2	9.8	
	Storage Tank	8.8	9.9	9.6	9.0	9.3	7.2	8.8	
	Distribution System	8.1	9.4	9.6	9.1	8.6	8.1	7.6	
Turbidity (NTU)	Raw	0.26	0.32	0.35	0.29	2.4	1.5	3.6	
	Launder	1.60	0.58	1.40	0.91	0.64	0.52	0.44	
	Filter Top	1.40	0.76	0.87	0.51	0.33	0.39	0.29	
	Filter Bottom	0.24	0.20	0.18	0.26	0.26	0.11	0.16	
	Clear Well	0.21	0.16	0.15	0.21	0.18	0.14	0.13	
	Storage Tank	0.15	0.11	0.11	0.11	0.14	0.10	0.14	
	Distribution System	0.16	0.18	0.16	0.17	0.21	0.16	0.23	
Alkalinity (mg/L as $CaCO_3$)	Raw	288	298	296	292	296	300	300	
	Launder	88	244	108	132	112	116	130	
	Filter Top	28	44	48	24	12	16	36	
	Filter Bottom	28	50	48	30	8	8	32	
	Clear Well	28	56	36	34	12	8	28	
	Storage Tank	28	42	22	14	16	8	12	
	Distribution System	28	34	38	16	12	8	12	

Table 27. Cost Comparisons for the Regular and High-Lime Treatment Processes at Acme Water Treatment Plant.

Parameter	Regular Treatment ($/1000 gallons)	Opt. Treatment ($/1000 gallons)	Increment ($/1000 gallons)
Chemicals	0.1211	0.1864	0.0653
Labor	0.1218	0.1218	0
Electrical	0.1069	0.1069	0
Total	0.3498	0.4151	0.0653
THM's	440 μg/L	222 μg/L	218 μg/L

The organic parameters (DOC, THM, TOX, color) show the beneficial effect of high pH treatment during the AID plant study. The DOC removal due to softening was 43% at the beginning of the plant test and was increased approximately 7% for the duration of optimization. The percent of DOC removal during the test is shown in Figure 13. This increase represents an additional 1 mg/L DOC and is similar to the bench scale results.

The THM's measured in the plant storage were 440 µg/L at test initiation. THM's were decreased approximately 200 µg/L ranging from 286 to 188 µg/L in the reservoir and from 387 to 175 µg/L in the distribution system. A definite time lag in the THM drop in the distribution system is apparent in Table 26. The increased % THM removed is shown in Figure 13 for the study and is close to 60% for optimization. This drop is caused by increased organic removal during softening and a decreased chlorination dose. The data in Table 26 shows an adequate Cl_2 residual was maintained in the distribution system once the Cl_2 dose was increased to 14-mg/L prior to filtration and 3 mg/L in the plant reservoir. The decrease in THM formation during the plant test was similar to that obtained in the jar test which approaches 50% relative to plant samples. The chlorine dose, however, was about 50% low. The chlorination curves had indicated a dose of 8 to 10-mg/L would be adequate whereas 17-mg/L was finally utilized. This difference was probably due to the difference in demand between a glass laboratory reactor and the actual demand in piping, tanks, etc.

The TOX taken from the storage tank were 660 µg/L and were 1.5 times the THM at the same location. This ratio is convenient, but it must be remembered that the ratio is µg/L Cl to µg/L THM. On an equivalent carbon mass, the TOX would be even larger in relation to the THM. For example, a chloroform molecule is 89% Cl or 89% of the THM is TOX. When more than single chain carbons are added, the weight difference between TOX and THM increases. As shown in Figure 13, the increased TOX removal is from 40% to 20% with the lower number being observed on the last day of plant testing. The average decrease in TOX was 190 µg/L or 29%.

6.1.1.7 AID Cost

Attached in Table 27 is a summary of the cost comparison between the regular and optimized treatment processes as practiced at AID. As can be seen, the difference in cost is attributable to an increased cost in chemicals for the optimized process. This increase of 6.53 cents per 1000 gallons is due to a combination of increased lime dosages, the use of sulfuric acid for pH control and a slight lowering of the chlorine dosage.

During the High-Lime Treatment period a 50 percent drop was found to occur in THM's. This represents an average drop of 218 u/L THM. THM reductions are based on the storage tank sample with the August 1 sample representing the baseline and an average of the storage tank values for the period August 4 through August 7. These dates were chosen to allow a "flushing" of the storage tanks and give a reasonable assurance that the water being analyzed was produced by the optimized treatment. The THM's at the plant storage tank was 440 µg/L on 8/1/84. The 8/4/84 to 8/7/84 THM's

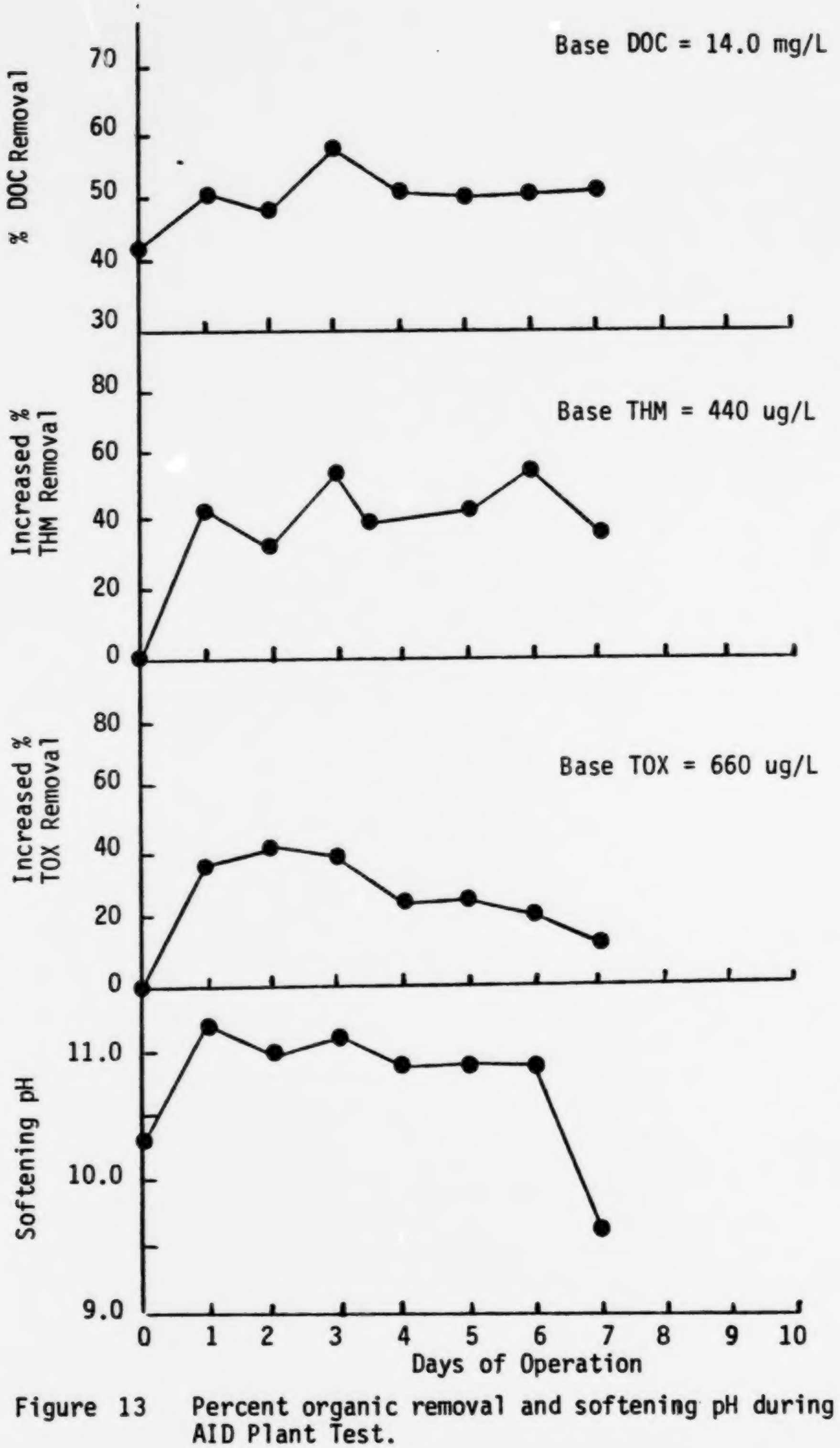


Figure 13 Percent organic removal and softening pH during AID Plant Test.

averaged 222 µg/L at the plant storage tank. The unit cost of water treatment at AID was \$0.35/1000 gal which increased 19% to \$0.42/1000 gal due to the optimized lime-softening treatment.

The plant testing at AID did show a general improvement in organic water quality due to increased softening pH and decreased Cl_2 dose. As the plant and laboratory data show increasing the reaction pH from 10.3 to 11.3 will decrease AID organic removal 10% to 20%. Once organic removal is increased, Cl_2 demand is decreased. The reduction in Cl_2 dose will reduce THM's as will a reduction in DOC at a constant Cl_2 dose. When both the DOC and Cl_2 dose are reduced, THM's should be reduced from 20% to 50%. The cost of such treatment changes should increase O&M costs at AID by 19%.

6.1.2 VOG

6.1.2.1 Introduction

The Village of Golf Utilities (VOG) is located in Palm Beach County, Florida approximately 10 miles south of West Palm Beach. The utility serves a predominantly residential population with significant seasonal variations in population and water use. Currently the utility serves approximately 1,000 connections with an approximate population of 2,500.

The raw water supply is supplied by three wells which tap a shallow aquifer locally known as the Turnpike Aquifer. This sand and shell aquifer is recharged via rainfall and surface water infiltration. The raw water characteristics are given in Table 28.

The plant utilizes the lime-soda ash method of softening. The treatment train may be seen in Figure 14. Basically unaerated (the aeration unit was not used) raw water is treated with lime, soda ash, a coagulant and settling aid (Polyhall 295) and chlorine in an accelator unit which is rated at 1 MGD. The settled water is then filtered through a multi-media filter and stored in above-ground storage tanks. The filtered water is routinely chlorinated again within the storage facilities. The current typical chemical dosages and finished water characteristics may be seen in Tables 29 and 30 respectively.

6.1.2.2 Jar Test

The first step in optimizing the lime-soda ash process for organics removal involved the application of increasing doses of lime with the subsequent raising of the softening pH. In the lime-soda process a floc of $CaCO_3$ and $Mg(OH)_2$ (at pH > 11) is formed. Organic compounds such as humic and fulvic acids are removed to varying degrees by this process. As the pH is increased solubility of calcium as $CaCO_3$ and magnesium as the hydroxide is reduced and a greater amount of floc is formed. This higher pH should also theoretically remove a greater amount of organic material.

Raw water was obtained from the VOG treatment plant, transported back to UCF and stored at 4°C to retard chemical and biological alteration of the raw water. Jar tests were conducted using one-liter aliquots with

Table 28. Village of Golf Raw Water Quality

Parameter	Value
pH	7.40
DOC	14.20 mg/L as C
TOX	50 µg/L as Cl
THM	15 µg/L
Color	52 cpu
TDS	340 mg/L
Total Hardness	240 mg/L as $CaCO_3$
Calcium Hardness	220 mg/L as $CaCO_3$
Conductivity	0.420 mmhos

Table 29. Village of Golf Typical Water Treatment Chemical Doses

Chemical	Dose
Lime	175 mg/L as $Ca(OH)_2$
Soda Ash	75 mg/L as Na_2CO_3
Polymer	0.7 mg/L Polyhall 295
Chlorine	26 mg/L as Cl_2

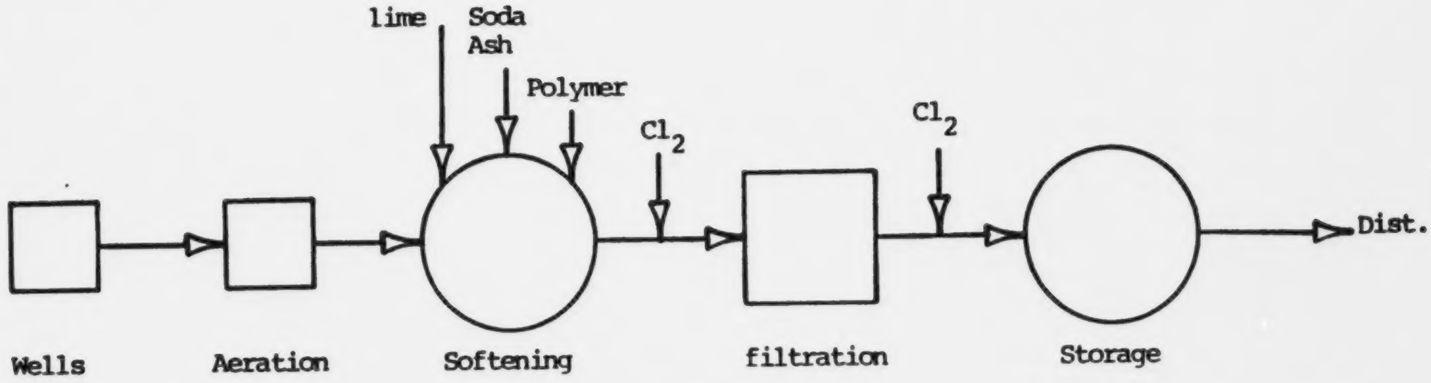

lime
Soda Ash
Polymer
Cl_2
Cl_2
Dist.
Wells
Aeration
Softening
filtration
Storage

Figure 14 Village of Golf Water System.

Table 30. Village of Golf Finished Water Quality

Parameter	Value
pH	8.80
DOC	13.50 mg/L as C
TOX	1350 μg/L as Cl
THM	600 μg/L
Color	10.0 cpu
TDS	230 mg/L
Total Hardness	50 mg/L as $CaCO_3$
Calcium Hardness	50 mg/L as $CaCO_3$
Conductivity	0.320 mmhos

rapid and slow mix times of one and twenty minutes respectively and five minutes of settling. Samples were taken from a depth of 10 cm which corresponded to the plant surface loading rate. Chemicals were applied in the following order: lime, soda-ash, polymer. Table A-2 and Figure 15 give the data for the jar tests. From the data the following trends may be noted:

1. The organic parameters color, DOC and THM's show a general decreasing trend with increasing pH. Color and DOC reach their minimum at pH 11.5 while THM's reaches a minimum at pH 10.6.
2. Conductivity (and thus TDS) exhibits a large increase between pH 11.0 and pH 11.5.
3. TOXFP exhibits a pattern similar to the other organic parameters.

Based on the above data and trends, it was decided to use pH 11.3 for further testing. It was judged that this pH would allow the advantage of good organics removal noted at the high pH values without the excessive TDS values found at the highest tested pH (11.5). It was also noted that this pH could be achieved without excessive chemical doses.

6.1.2.3 Coagulant Aid

Subsequent to the Process pH Study, three coagulants were examined for organics removal. The three chosen coagulants were commercial alum and two polyaluminum chloride solutions PACl 190 and PACl 100S.

The jar tests were conducted using one liter aliquots with the coagulant added first then followed by the normal chemical sequence. Doses equal to 0.8, 1.6, 2.4 and 4.0-mg/L as Al were utilized. The data for DOC, color, THMFP, and Al residuals can be found in Table A-3 and Figures 16, 17, 18, and 19. It should be noted the control jar was treated at pH 11.3 without any coagulants.

The data obtained in this study does not indicate the clear superiority of any one coagulant aid. However certain trends as in the AID study can be seen:

1. All coagulant aids cause a decrease in DOC when compared to the control jar. Alum gives the lowest average DOC value.

2. All coagulant aids demonstrated no ability to reduce trace color. The one exception was alum at 1.6-mg/L as Al.

3. All the coagulant aids cause a slight decrease in THM's; however no consistent trend is noted among the three coagulants.

4. Alum clearly gives a lower aluminum residual at all aluminum doses.

Based on the above data and trends it was decided to utilize alum at an equivalent aluminum dose of 0.8-mg/L (approximately 8.5-mg/L as Alum). This coagulant and dosage were chosen for the following reasons at AID:

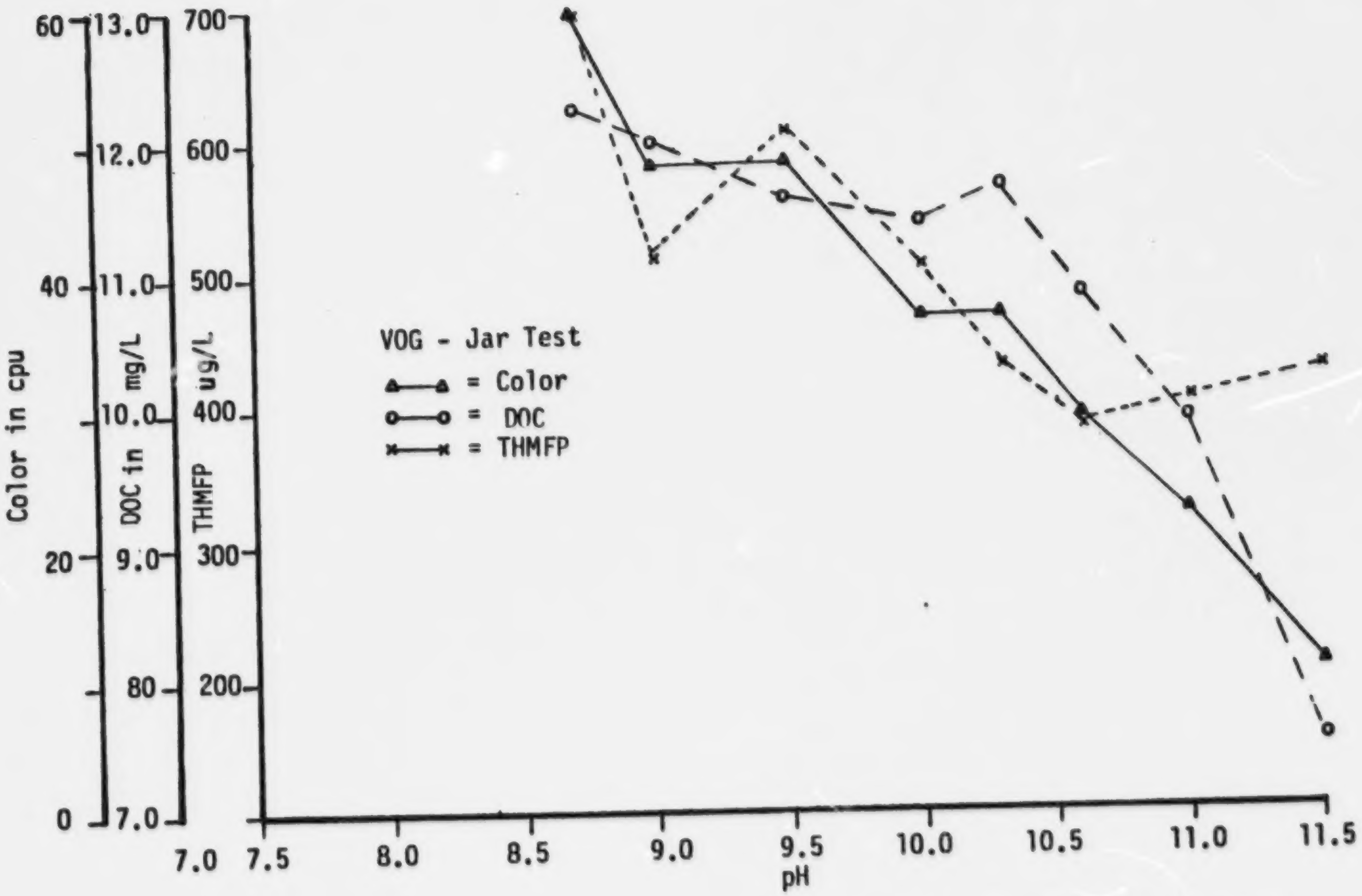


Figure 15 Village of Golf Process pH Optimization Study - Color, DOC and THMFP vs pH

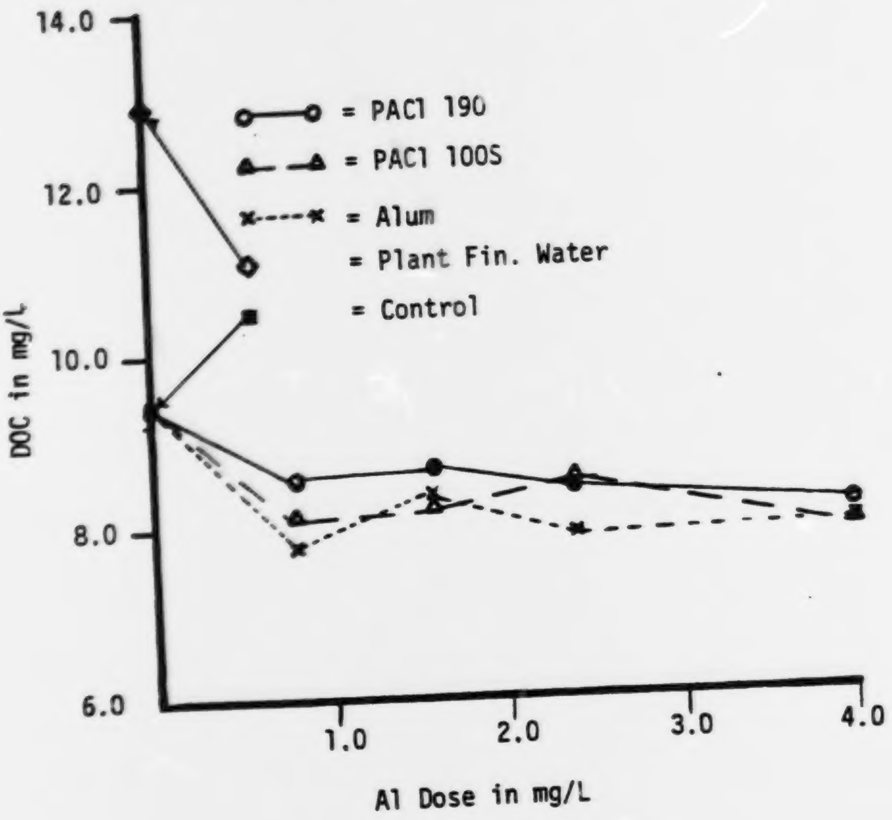


Figure 16 Village of Golf - DOC vs Al Dose

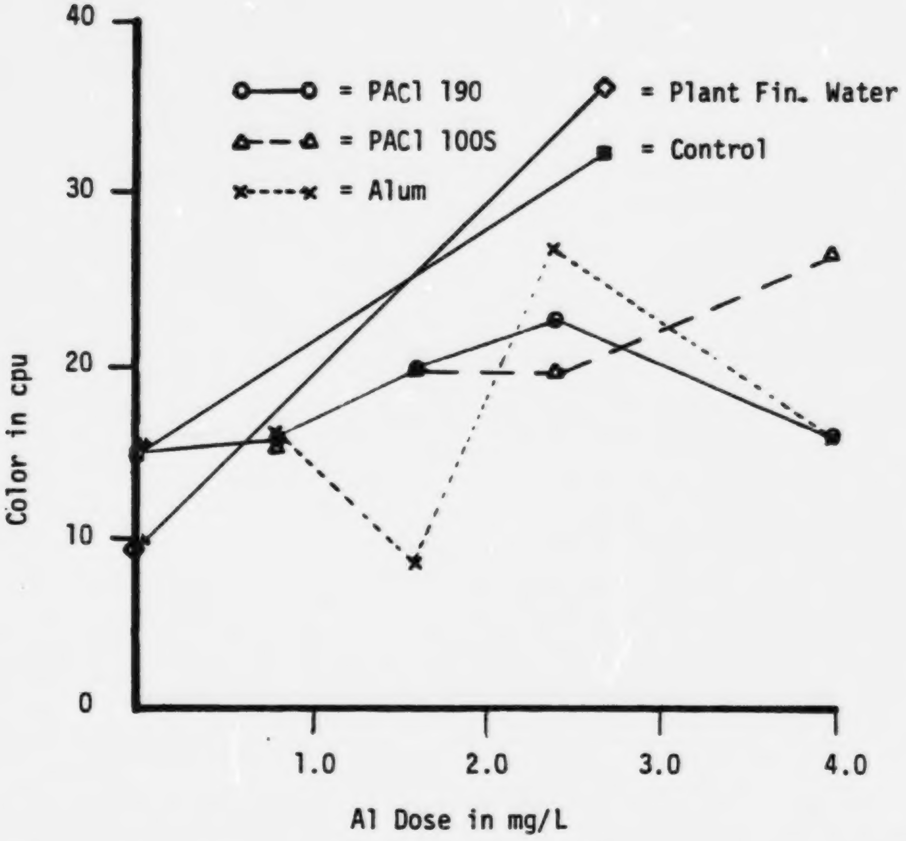


Figure 17 Village of Golf - Color vs Al Dose.

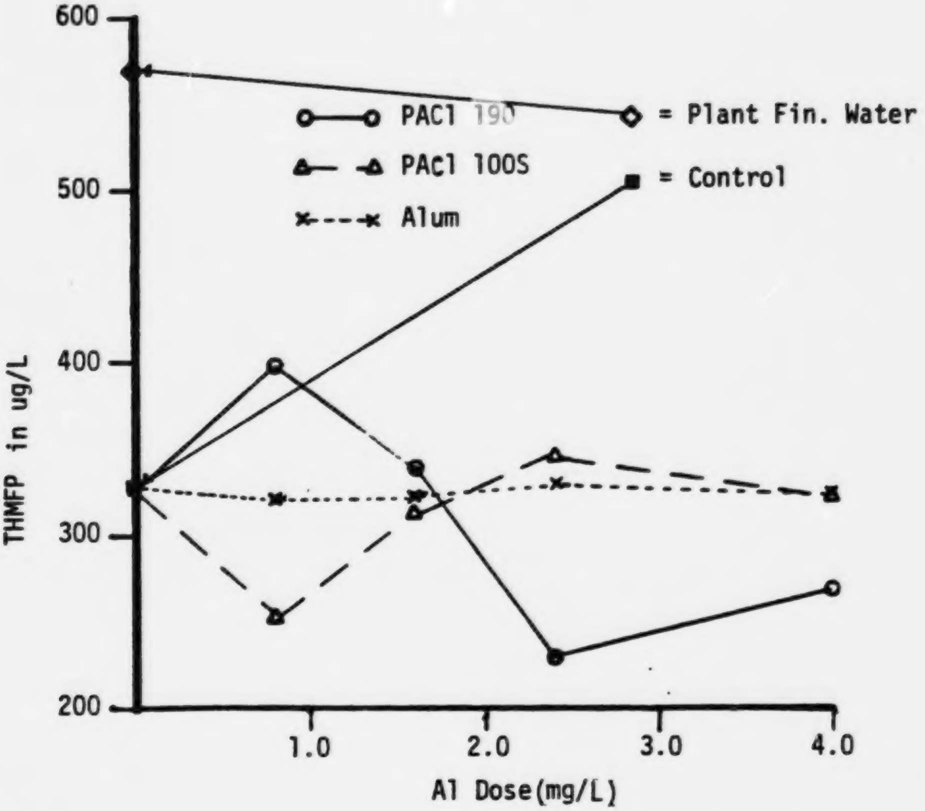


Figure 18 Village of Golf - THMFP vs Al Dose THM values are 48 hr. FP at 20° C, Initial Cl_2 dose was 26 ppm.

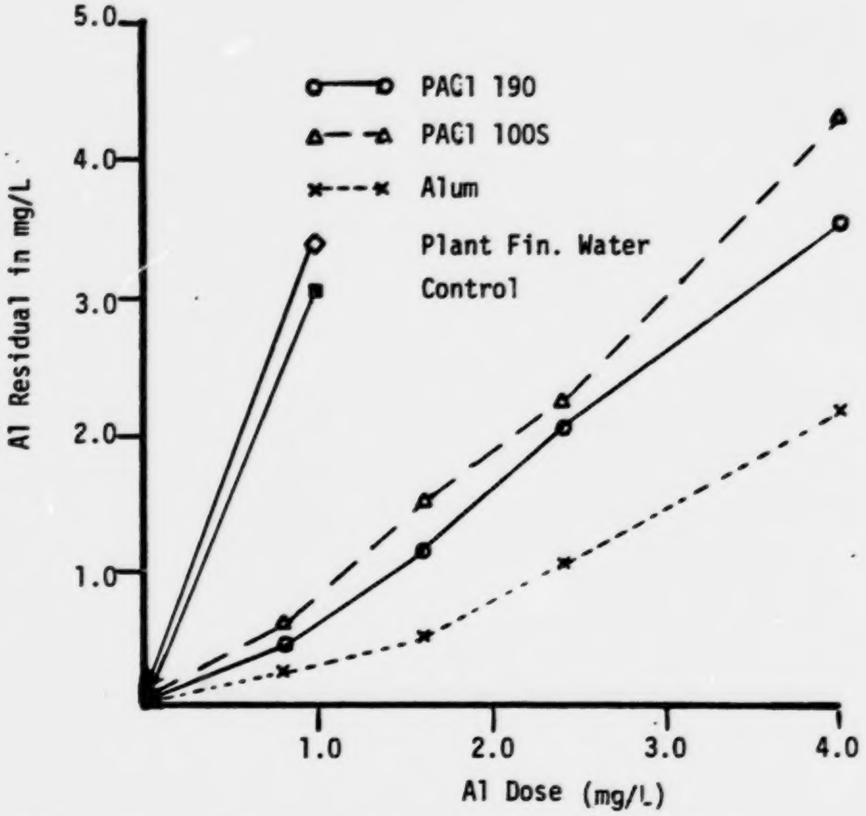


Figure 19 Village of Golf - Al Residual vs Al Dose

1. Alum generally gave an improvement in organics removal. However an increase in dosage did not in general improve the organics removal.

2. Alum at this dose gives the lowest aluminum residual. This is significant if MCLs are set for aluminum in the future.

3. Alum has a significant history of use in the U.S. Water Treatment Industry and enjoys an advantage in cost and availability.

4. Visual observation of Jar Tests indicates that Alum as the other coagulant aid improves floc settling.

6.1.2.4 Chlorine Dissipation and THM Formation

During this phase comparisons were made between the water produced using the plant chemical dosages and waters produced using the UCF designed chemical dosages. Several studies were conducted to compare the waters and these are given below:

1. Thirty Minute Cl_2-Induced Color Removal - This study compared the color remaining after 30 minutes contact using various chlorine doses for the two waters. As can be seen in Figure 20 and Table A-4 a dose of 5.0-mg/L Cl_2 was sufficient to reduce the color of the UCF water to the 15 cpu limit while a 15.0-mg/L dose was necessary for the plant water. Thus a significant advantage of 10.0-mg/L Cl_2 was noted.

2. Chlorine Dissipation Curves - This study served two different purposes. First various chlorine doses were applied to the UCF water to determine a dose that would allow a free chlorine residual of at least 0.2-mg/L at the end of 96 hours (4 days). Second a comparison could be made between these dosages and the current plant dosage of 26-mg/L Cl_2. Table A-5 and Figure 21 give the data. As can be seen, a chlorine dose of 20.0-mg/L will allow the UCF water to retain a significant residual at the end of 96 hours (4 days). Also, from the 24 hour mark to the 96 hour mark, this curve is quite similar to the plant dissipation curve. Since the 20-mg/L Cl_2 dose gave the UCF water a significant residual at the end of 96 hours, it was decided to utilize a 15-mg/L initial Cl_2 dose for all future tests.

3. THM Formation Curve - This phase compared the formation of THM with time for the two waters. It is important to note that the incubation of these samples was conducted at 30°C. This temperature is approximately 10°C higher than normally encountered in the system. In this manner a "worst-case" formation could be determined. Tables A-6, A-7, 31, and Figure 22 give the results of the study. The water temperature of finished water in Florida varies from 20°C to 30°C. Since VOG has high THM's as indicated by previous testing, the incubation temperature was maximized at 30°C.

It can be seen that the optimized water has a THM's that is approximately one-half the plant formation rate. Such values can be attributed to

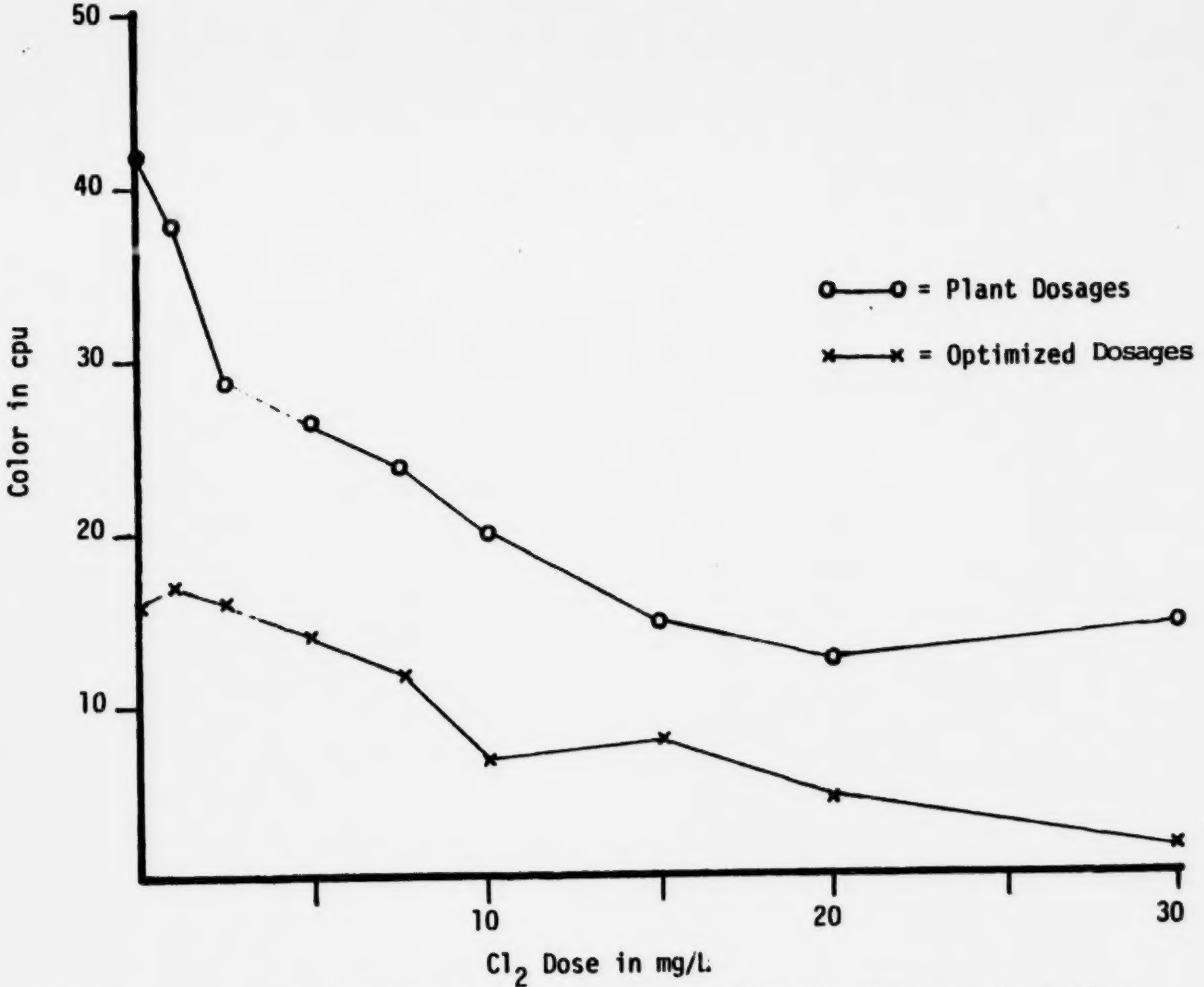


Figure 20 Village of Golf - Color vs Cl_2 Dose after 30 minutes contact time.

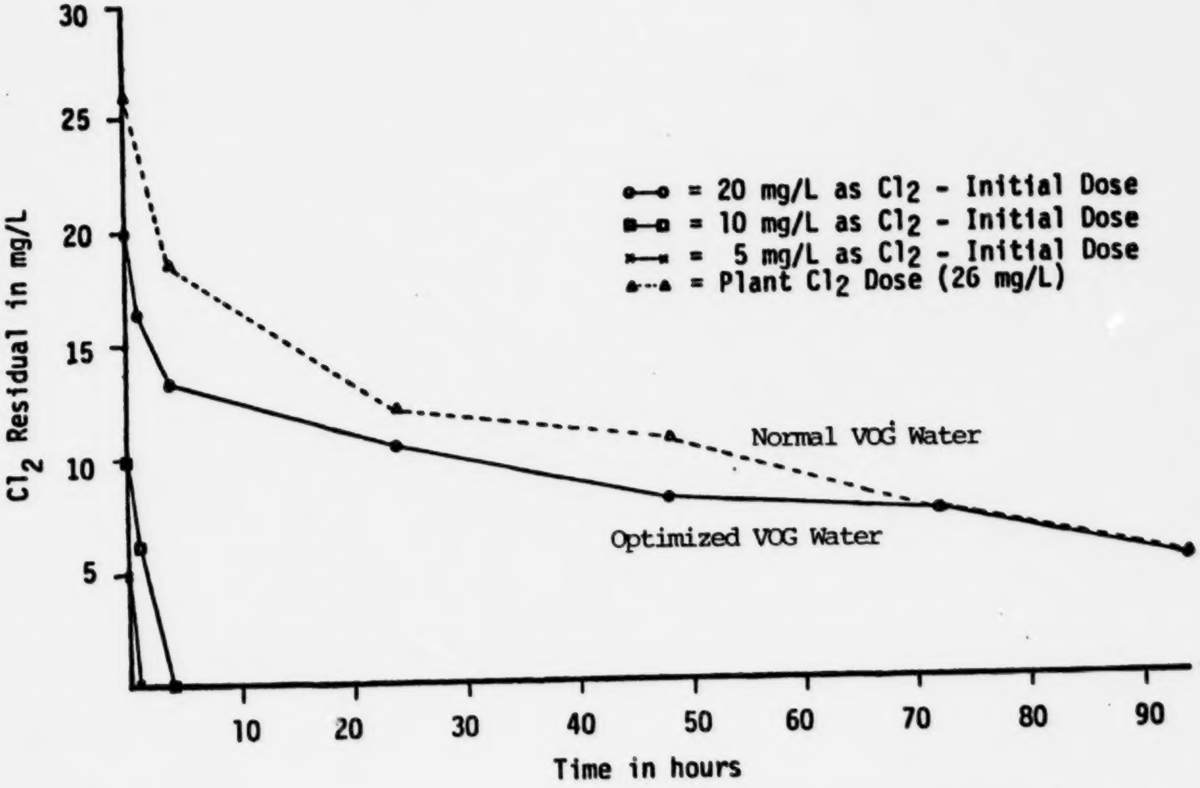


Figure 21 Village of Golf Chlorine Dissipation Curve at 30° C

Table 31. Village of Golf - Comparison of THM Formation Curves
(Plant vs UCF Dosages)

Time in hours	Plant THM in µg/L	UCF THM in µg/L	UCF Decrease	
			in µg/L	%
Instant	104	34	70	67
1	192	107	85	44
24	857	402	455	53
48	953	534	419	44
72	1060	642	418	39
96	1112			

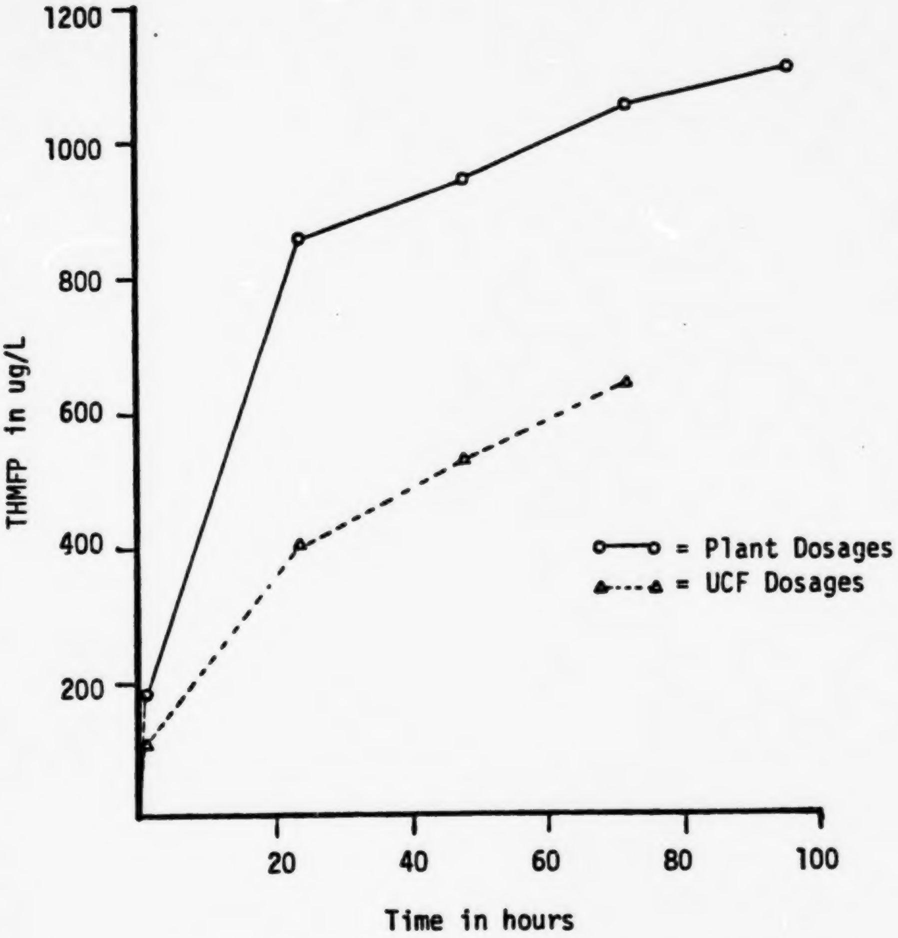


Figure 22 Village of Golf - THM Formation Curve Plant Dosages vs UCF Dosages at pH 8.8 and 30^{o} C

the lowered DOC and color values in conjunction with the lower chlorine dose (15-mg/L as compared to 26-mg/L).

6.1.2.5 Recommendations

Based on the VOG bench scale studies, the following recommendations were made:

1. A five-day continuous plant scale test was conducted using the optimum chemical doses. These doses are given in Table 32 and are compared to the present dosages.

2. The chlorination point was moved to a point after the water is coagulated but prior to the filter unit. For algal control in the contact tank a small pre-treatment chlorine dose could be utilized.

3. Four sampling points were established: the raw water, the settled water prior to the filters, the clear well, and in the distribution system a suitable distance from the plant. Standard parameters measured included DOC, color, THM, conductivity, total hardness, pH, chlorine residual, and aluminum residual. Three samples were taken each day.

6.1.2.6 VOG Plant Optimization

The VOG plant test was conducted from August 6, 1985, to August 11, 1985. This is a small plant that serves a seasonal population that is very low during the summer months. Although the exact service population was not known during the plant test, the VOG personnel estimated that less than 200 of the 2500 population were present. Consequently the distribution system samples are probably not representative of the water produced during the plant test. To offset the lack of consumer use, two samples were taken from the 400,000 gallon storage tank. One was immediately quenched and one was quenched after 48 hours. However during the plant test, the average hours of operation and production were 8.4 hrs. and 228,000 gal. The plant was rated at 1.0 MGD, produced at a rate of 0.653 MGD which meant the SLR was 0.65 gpd/ft^2 and therefore was not under hydraulic stress.

The chemical feed system is unusual in that lime and soda ash are batch mixed simultaneously in the same tank. The VOG plant utilizes soda ash to reduce noncarbonate hardness feeding an average of 5.1 meq/L $Ca(OH)_2$ and 2.2 meq/L Na_2CO_3. The raw water hardness and alkalinity are 5.0 meq/L and 4.4 meq/L. The product hardness is 1.0 meq/L calcium and 0.2 meq/L magnesium. Some $CaCO_3$ precipitation probably occurs in the batching process but the product hardness does reflect the removal of noncarbonate hardness. This is somewhat surprising since all of the added alkalinity is available for precipitation in the batching process.

The chemical feed is not continuous. The feed pump runs 45 seconds every 30 minutes and doses enough chemicals to treat 30 minutes of raw water. This obviously causes pH variations, 11.3 to 8.3, on a cyclical basis. However, plant personnel (3 in total) have responsibility for a

Table 32. Village of Golf - Comparison of Plant and UCF Chemical Doses

Chemical	Plant	UCF	Change in mg/L	Change in lbs/MG
Lime as $Ca(OH)_2$	175 mg/L	350 mg/L	+175 mg/L	1460
Alum as Al	0	0.8 mg/L	+0.8 mg/L	6.67
Soda Ash as Na_2CO_3	75 mg/L	75 mg/L	0	0
Polymer as Polyhall 295	0.7 mg/L	0.7 mg/L	0	0
Chlorine as Cl_2	26 mg/L15	mg/L	-11.0 mg/L	-91.7

wastewater plant, the drinking water plant, the distribution system, complaint investigation, meter readings, etc. and obviously do not have spare time to mix chemicals.

The normal plant process has a Cl_2 dose of 20 mg/L prior to filtration and another Cl_2 of 6 mg/L prior to storage. The pH onto the filters is usually high (9.2) and will precipitate $CaCO_3$. The high pH is necessary to offset the pH drop caused by chlorination. This method of final pH control would only work at a plant for a water, like VOG, which has removed all the carbonate buffering. The modifications for plant optimization were like AID in that the Cl_2 dose was lowered from 20-mg/L to an average of 13-mg/L prior to filtration and 6-mg/L prior to storage. Sulfuric acid was fed to the settled water to offset the lower Cl_2 dose. VOG plant personnel commented that the lower filtration pH reduced the solids formation on the filters and eased filter cleaning at the conclusion of the plant test.

The results from the VOG plant test are presented in Table 33. The DOC, color, and THM's were measured at the raw, launder, filter top and bottom, storage tank at 0 and 48 hours, and the distribution system. TOX was only measured at the storage tank and the distribution system. The inorganic parameters; pH, alkalinity, hardness, and Cl_2 residual were measured at the raw, launder, filter tap and plant tap.

The raw DOC averaged 19.2-mg/L for the test and was reduced to 12.2-mg/L (37%) through softening and to 12.0-mg/L by filtration. This DOC reduction was generally expected as jar tests had indicated a 40% DOC reduction above pH 11.0 was possible. However, the well utilized for the test has a high DOC compared to the well utilized for the jar test. As a result, the final DOC was 2 to 4.5-mg/L higher than the bench work.

The raw color averaged 61 cpu and was reduced to 18 cpu (71% reduction) through softening. A final color of 6 cpu was produced after filtration and chlorination, 13-mg/L prior to filtration and 6-mg/L in the plant reservoir. The increased color reduction from the higher pH was not noticeable because the plant normally chlorinated in the reactor which appeared darker during the test.

The THM at the storage tank and distribution system averaged 414 μg/L overall and 257 μg/L on 8/10 and 8/11. The storage tank THM's returned to 692 μg/L on 8/13 which is due to a return to normal operation. At worst the THM reduction was 40%, at best 63%. The average THM's in the 48-hour tank sample and the distribution system were 475 μg/L and 715 μg/L. The 48-hour storage tank samples represent a 22% reduction and the distribution samples have questionable value because of low demand. However observing the THM data in Table 40 definitely shows a significant drop in THM.

The TOX samples show a more stable trend than THM's. Looking at the last three days again a general drop is observed. The storage tank and reservoir samples averaged 517 and 619 μg/L respectively during this period. This represents a TOX reduction of 20% to 40% over normal plant operation.

Table 33. VOG Plant Optimization Test - Inorganic Parameter Values

Parameter	Sample Location	8/6	8/7	8/8	8/9	8/10	8/11	8/12	8/13
DOC (in mg/L as C)	Raw	17.50	20.35	19.64	29.65	19.36	18.34		19.43
	Launder	11.32	13.46	11.55	12.10	11.93	12.67		13.08
	Filter Top	11.39	11.44	11.76	11.61	12.05	11.36		13.85
	Filter Bottom	11.38	12.52	11.44	12.23	12.23	12.30		13.48
	Storage Tank (0)	12.85	12.26	12.22	12.23	12.04	12.47		11.14
	Storage Tank (48)	12.76	12.65		12.74	11.96	11.82		11.77
	Distribution System	12.96	10.79		11.36	11.43	11.23		10.90
Color (in cpu)	Raw	57	64	63	65	64	53		63
	Launder	15	22	17	20	15	18		23
	Filter Top	11	22	21	22	21	17		14
	Filter Bottom	9	23	17	17	10	18		18
	Storage Tank (0)	13	15	16	13	14	9		9
	Storage Tank (48)	14	12	10	12	11	11		10
	Distribution System	9	6	6	9	7	7		8
THM (in µg/L)	Raw	0	0	0	0	0	0		0
	Launder	69	0	0	25	-	0		0
	Filter Top	180	219	37	37	27	27		37
	Filter Bottom	254	219	154	146	47	90		579
	Storage Tank (0)	475	517	613	364	239	275		692
	Storage Tank (48)	442	720	691	634	264	314		608
	Distribution System	668	1125	--	596	654	534		748
TOX (µg/L as Cl)	Storage Tank (0)	794	586	476	498	475	577		623
	Distribution System	1161	1101	--	482	558	817		1140

(continued)

Table 33 (continued). Inorganic Plant Test Data

Parameter	Sample Location	8/6	8/7	8/8	8/9	8/10	8/11	8/12	8/13
	Raw	7.6	-	7.2	-	-	74		
	Launder	9.5	11.0	11.0	11.0	10.9	10.9		
	Filter Top	9.0	9.5	9.0	10.0	9.5	9.5		
pH	Finished	8.4	8.5	8.0	9.1	8.6	9.3	8.6	8.7
Alkalinity	Raw	220	-	215	-	-	220		
	Launder	150	140	170	172	168	185		
(mg/L as	Filter Top	80	90	50	49	50	40		
$CaCO_3$)	Finished	52	57	44	36	45	48		
TH	Raw	245	-	252	-	-	246	-	
(mg/L as	Launder	90	100	110	140	120	120		
$CaCO_3$)	Filter Top	-	-	-	-	-	-		
	Finished	95	104	110	101	105	90		
Cl_2	Raw	-	-	-	-	-	-	-	
(mg/L as Cl_2)	Launder	23	17	10	14	14	14	13	18
	Filter Top	7	9	5	8	9	9	5	5
	Finished	1.9	1.9	1.5	2.5	1.0	1.4	1.4	1.0

The inorganic plant data is also included in Table 33. It shows what would be expected by increasing the softening pH. The finished hardness increased to about 100 mg/L. Final alkalinity was low, around 50 mg/L and a Cl_2 residual was maintained at the plant tap.

6.1.2.7 VOG Cost

The cost information supplied by the VOG management is detailed, but in a different format than most plants utilize. The cost of plant operation for VOG is given in Table 34 as supplied by VOG with the additional chemical costs for the seven days of optimized treatment. This additional cost does not include the cost of the H_2SO_4 which was supplied by the project. The H_2SO_4 can be purchased for $1.00/gallon in bulk, although for this project it was purchased at $1.285/gallon in a limited quantity (150 gallons). The cost of the optimized treatment was determined by attributing the extra VOG, project chemical cost plus the cost of 57 gallons of H_2SO_4 to the optimized process. The THM's at VOG were reduced by approximately 50%, from 581 to 289, which increased the chemical cost by $0.23/1000 gallons for the seven days of optimized operation. The $1.40/1000 gal cost for August 1984 represents an average cost for the month and does not reflect a cost of continuous optimization.

The normal THM production at VOG has been measured at 450 and 750 μg/L by other laboratories. UCF laboratories have measured VOG THM's from 600 to almost 1200 μg/L during plant tests in the distribution system. Using the 692 μg/L THM measurement at the storage tank two days after the study ended as a base and the overall weeks average of 417 μg/L for a 40% THM reduction is conservative and is presented with the 63% best case reduction of 257 ug/l. As shown on Table 35, the cost was increased from $1.33/1000 gal to $1.55/1000 gal which represents a 17% cost increase for THM reduction ranging from 36% to 64% or approximately 50%. The variability in the VOG data is due to seasonal occupation and differences in operation. However this data as does the AID data shows THM reduction at a lime-softening plant can be obtained by increasing softening pH and lowering Cl_2 dose.

6.1.3 Venice

6.1.3.1 Introduction

The Venice plant optimization was the third plant that was optimized for organic removal and had a lime-softening process. No laboratory jar tests were conducted using the Venice water. There were two reasons for this. First project personnel has previous experience with three Florida softening plants, two on this project and one in Melbourne where an increase in softening pH had decreased THM's in the distribution system. Second, Venice was originally an alternate site and project personnel time was being utilized on other project work. Consequently, a meeting between the project director and Venice personnel occurred in which a plant optimization was proposed and agreed to, based on the results of other plant and bench work.

Table 34. Operating Expenses for the Village of Golf, FL Municipal Water Treatment Facility

Time	1983-84	Aug 1984
Expense	$142,125	$8,843
Million Gals Treated	106.520	6,336
Cost per 1000 Gallons	$1.33	$1.40

Note: Above EXPENSE includes all current (out-of-pocket) operating costs. It does not include depreciation, renewal and replacement and reserve.

Table 35. Village of Golf Cost Comparisons for the Regular and Optimized Water Treatment Plant

	Treatment Regular	Treatment ($/1000 gal) Optimum	Increment
Total	1.33	1.55	+0.23
THM µg/L	692	441 to 257	-251 to -435

The Venice operation consisted of both lime softening and RO. The lime-softening plant took raw water from wells that has an exceptionally high hardness (700-mg/L $CaCO_3$ TH, 170-mg/L $CaCO_3$ MgH), moderate alkalinity and color (280-mg/L $CaCO_3$ and 35 cpu). The typical DOC in the raw water for softening was 15 to 17-mg/L. The wells for the RO plant typically produced 60% to 70% of Venice's 1.8 ±0.2 MGD. This ratio changed to 50% RO and 50% lime softening because of a fouling problem in the RO membranes which reduced their flux and production. The RO production was always maximized at Venice because a lower hardness could be produced. This operating philosophy also minimized THM's because of the 90% organic rejection of the RO membranes.

The plan for optimization was similar to that at AID and VOG. First the softening pH would be raised to 11.3. This would accomplish even more organic removal at Venice than either VOG or AID because the high magnesium in the raw water would precipitate and would remove additional organics. Second chlorination would be discontinued in the softener and be increased slightly after blending the RO and lime softened effluents in plant storage. Venice already was adding alum to the softening reactor to improve settling and color removal. Their process, as shown in Figure 23, involved aeration, softening, prechlorination, filtration, RO blending, postchlorination, and storage. Our changes were to reduce chlorination prior to filtration with acid and increase softening pH while maintaining maximum RO production. As at VOG and AID, chlorination was used to reduce pH prior to filtration as well as for trace-color removal and disinfection. Reducing the pH with acid reduced the required Cl_2 dose. The test was conducted from 11/30/84 to 12/17/84. However due to equipment failure during start up, acid feed was inconsistent, but all other Cl_2 feeds and pH changes were implemented.

6.1.3.2 Venice Plant Optimization

Only organic water quality is reported in Table 36 for Venice as no appreciable change in inorganic water quality took place. Typically the final (blended) hardness was 120-mg/L $CaCO_3$. Alkalinity was 50-mg/L and water was stable. DOC reduction in the RO plant averaged 85% from 3.6-mg/L in the raw to 0.5-mg/L in the RO product. The DOC reduction during lime softening averaged 46% from 16.0-mg/L to 8.5-mg/L which is representative of a lime-softening process operating at pH 11.0. However, this removal could be increased to 60% if more magnesium was precipitated, i.e., a higher pH. The blended DOC was 4.5 on average which reflects the equal 0.9 MGD production by the RO and lime-softening plants.

The THM's were measured at four points throughout the distribution system. Previously THM's in the Venice distribution system had ranged from 45 µg/L to 379 µg/L with the higher values being attributed to increased production from the lime-softening plant and problems with chlorination control in the distribution system. The THM average in the Venice distribution system is shown in Figure 24 for the duration of the plant study. THM's were 150 µg/L for the first day and averaged 85 µg/L for the nine samples taken during the remainder of the study for an average 37% decrease. THM's ranged from 70 to 125 µg/L during this period. Cl_2

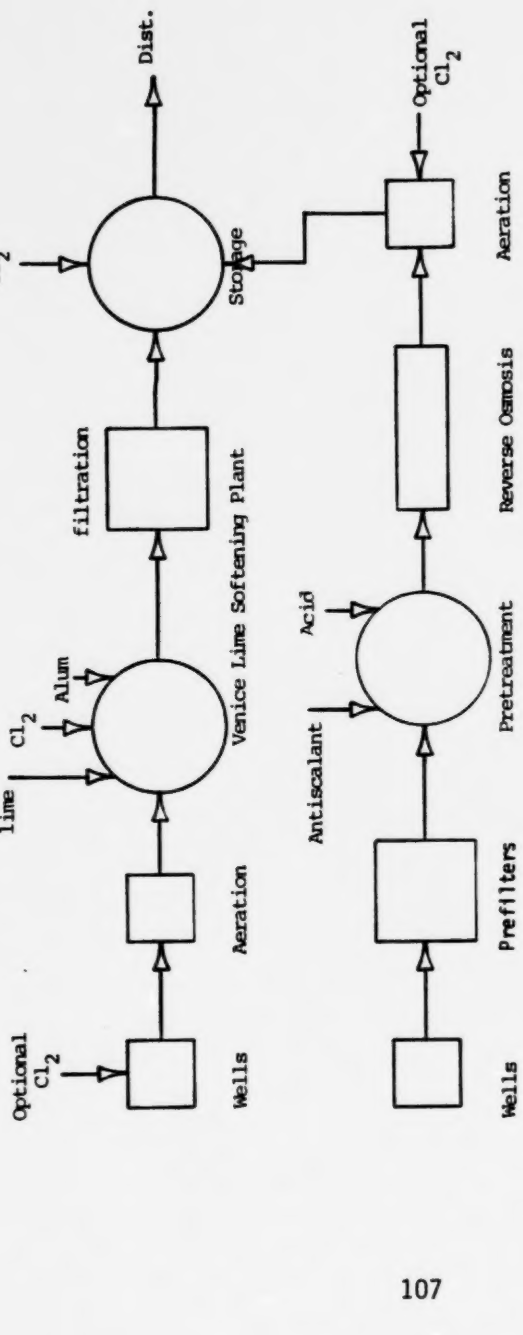

Optional Cl2
Wells
Aeration
lime
Cl2
Alum
Venice Lime Softening Plant
filtration
Cl2
Storage
Dist.
Wells
Prefilters
Antiscalant
Acid
Pretreatment
Reverse Osmosis
Aeration
Optional Cl2

Figure 23 Venice Potable Water System.

Table 36. Water Quality for the Venice Plant Optimization Study
(November 30 - December 4, 1984)

Date	Location or Type	Chlorine Residual (mg/L as Cl_2)	DOC (mg/L as C)	TOX (µg/L as Cl)	THM (µg/L)	Color (cpu)
11/30/84	R. O. Raw	0.0	3.4	22	4	3
	L. S. Raw	0.0	16.9	17	1	95
	R. O. Product	0.0	0.6	5	2	0
	L. S. Product	0.0	10.0	41	11	8
	Clearwell	N.A.	4.6	206	135	1
	Carvel	N.A.	4.6	253	156	1
	City Hall	N.A.	3.5	218	150	2
	Inlet Circle	N.A.	4.3	259	164	3
12/02/84	R. O. Raw	0.0	3.4	N.A.	N.A.	4
	L. S. Raw	0.0	16.6	N.A.	N.A.	90
	R. O. Product	0.0	0.4	N.A.	N.A.	0
	L. S. Product	0.0	10.2	N.A.	N.A.	9
	Clearwell	0.5	4.6	183	59	3
	Carvel	0.2	4.7	160	57	4
	City Hall	0.0	4.7	223	66	3
	Inlet Circle	0.0	4.2	220	99	2
12/04/84	R. O. Raw	0.0	3.4	N.A.	N.A.	2
	L. S. Raw	0.0	17.5	N.A.	N.A.	100
	R. O. Product	0.0	0.4	N.A.	N.A.	0
	L. S. Product	0.0	9.8	N.A.	N.A.	8
	Clearwell	0.5	4.5	183	118	2
	Carvel	0.2	4.7	223	99	4
	City Hall	0.0	4.2	249	111	4
	Inlet Circle	0.0	4.1	197	107	4

(continued)

Table 36 (continued). Water Quality for the Venice Plant Optimization Study (December 6 - December 9, 1984)

Date	Location or Type	Chlorine Residual (mg/L as Cl_2)	DOC (mg/L as C)	TOX (μg/L as Cl)	THM (μg/L)	Color (cpu)
12/06/84	R. O. Raw	0.0	3.2	N.A.	N.A.	3
	L. S. Raw	0.0	17.4	N.A.	N.A.	95
	R. O. Product	0.0	0.4	N.A.	N.A.	1
	L. S. Product	0.0	9.4	N.A.	N.A.	10
	Clearwell	0.5	4.3	157	56	3
	Carvel	0.1	4.4	165	63	2
	City Hall	0.0	4.0	175	99	3
	Inlet Circle	0.0	4.0	194	117	4
12/07/84	R. O. Raw	0.0	3.4	N.A.	N.A.	5
	L. S. Raw	0.0	17.3	N.A.	N.A.	95
	R. O. Product	0.0	0.6	N.A.	N.A.	1
	L. S. Product	0.0	9.8	N.A.	N.A.	10
	Clearwell	0.3	4.9	149	77	4
	Carvel	0.0	4.7	194	69	3
	City Hall	0.0	4.0	179	115	3
	Inlet Circle	0.0	4.3	216	96	4
12/09/84	R. O. Raw	0.0	3.7	55	0	5
	L. S. Raw	0.0	15.2	60	0	49
	R. O. Product	0.0	0.9	25	0	0
	L. S. Product	0.0	8.5	29	0	11
	Clearwell	0.5	4.4	197	99	2
	Carvel	0.4	4.3	205	81	2
	City Hall	0.1	3.9	151	95	10
	Inlet Circle	0.1	3.9	208	136	4

(continued)

Table 36 (continued). Water Quality for the Venice Plant Optimization Study
(December 10 - December 15, 1984)

Date	Location or Type	Chlorine Residual (mg/L as Cl_2)	DOC (mg/L as C)	TOX (µg/L as Cl)	THM (µg/L)	Color (cpu)
12/10/84	R. O. Raw	0.0	6.90	N.A.	N.A.	4
	L. S. Raw	0.0	14.82	N.A.	N.A.	50
	R. O. Product	0.0	0.57	N.A.	N.A.	0
	L. S. Product	0.0	10.74	N.A.	N.A.	16
	Clearwell	0.5	4.91	227	69	3
	Carvel	0.4	5.05	200	81	3
	City Hall	0.1	4.22	192	103	4
	Inlet Circle	0.1	4.31	192	106	4
12/12/84	R. O. Raw	0.0	3.74	N.A.	N.A.	5
	L. S. Raw	0.0	14.60	N.A.	N.A.	44
	R. O. Product	0.0	0.63	N.A.	N.A.	0
	L. S. Product	0.0	8.93	N.A.	N.A.	12
	Clearwell	0.5	4.75	187	101	3
	Carvel	0.4	4.65	198	98	3
	City Hall	0.1	4.38	177	103	5
	Inlet Circle	0.1	4.23	224	160	4
12/15/84	R. O. Raw	0.0	4.0	N.A.	N.A.	8
	L. S. Raw	0.0	14.7	N.A.	N.A.	43
	R. O. Product	0.0	1.0	N.A.	N.A.	0
	L. S. Product	0.0	7.7	N.A.	N.A.	8
	Clearwell	1.2	3.6	197	73	1
	Carvel	1.0	4.2	188	73	1
	City Hall	0.1	4.1	158	61	3
	Inlet Circle	0.0	3.9	207	126	3

(continued)

Table 36 (continued). Water Quality for the Venice Plant Optimization Study
(December 17, 1984)

Date	Location or Type	Chlorine Residual (mg/L as Cl_2)	DOC (mg/L as C)	TOX (μg/L as Cl)	THM (μg/L)	Color (cpu)
12/17/84	R. O. Raw	0.0	3.7	64	0.0	12
	L. S. Raw	0.0	14.7	57	0.0	47
	R. O. Product	0.0	0.6	15	0.0	0
	L. S. Product	0.0	8.1	42	0.0	10
	Clearwell	0.5	4.4	193	80	2
	Carvel	0.4	4.5	210	95	2
	City Hall	0.2	4.3	179	87	5
	Inlet Circle	0.1	4.0	201	140	4

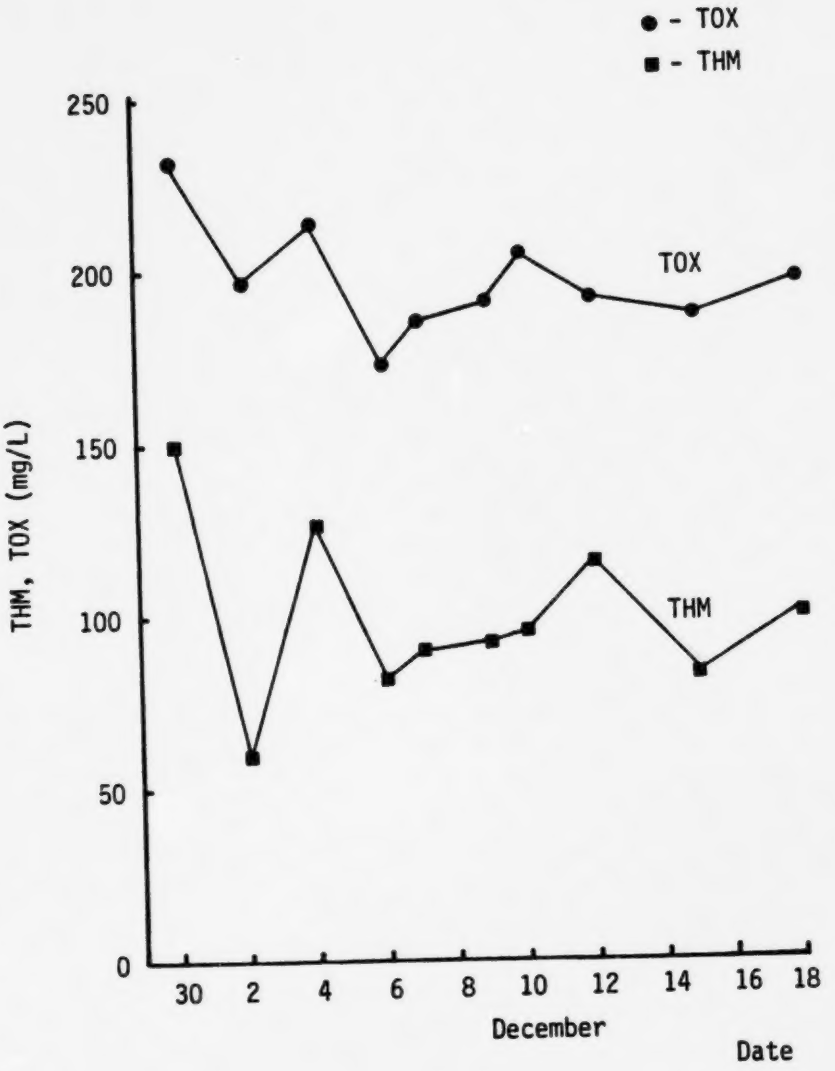


Figure 24 THM and TOX Averages during Venice Plant Tests in the Distribution System

residuals at the furthest points of the distribution system were not seen until ten days after the test began. After 10 days, a 0.1-mg/L residual was maintained.

The TOX averages are presented in Figure 24 with THM. Initially the TOX averaged 234 μg/L and was reduced to an average of 196 (18%) for the remainder of the study. The TOX curve closely follows the THM curve with time. The TOX/THM ratio moves from 1.6 to 2.8 and averaged 2.0 for the plant test. Color removal ranged from 75% to 85% with the higher removal corresponding to a higher color (95 vs. 45 cpu in the raw).

6.1.3.3 Venice Cost

The water quality from the Venice Optimization study is shown in Table 36. The costs of operation and optimization are shown in Tables 37 and 38. Basically the only changes in the normal method of operation of the lime-softening plant were (1) to move and reduce the chlorination point and feed and (2) to raise the softening pH. The costs from these two changes offset one another and did not represent a major cost increase for operation. The actual cost increase for these changes was \$0.02/1000 gal. It should be noted that the city of Venice actually had lower THM's than the 0.10-mg/L THM MCL during these studies. They took their regular quarterly sample at the same time as the optimization was occurring and their samples averaged 0.098-mg/L.

No acid was fed prior to filtration during optimization, however the pH should be reduced from the softening pH prior to filtration. There is a significant amount of magnesium, 150-mg/L as $CaCO_3$ in the raw water for the softening plant at Venice. Raising the softening pH to 11.0 should remove approximately 50% DOC and at Venice will reduce THMS from 30 to 50%. Acid will have to be fed to keep the filters from cementing. The most economical method of increasing softening pH is NaOH which costs \$0.085/lb and would increase chemical cost about \$0.025/1000 gal. The H_2SO_4 acid feed required to reduce the pH prior to filtration costs \$0.06/lb and will increase cost about \$0.025/1000 gal. It would also be advisable to feed a small amount of Na_2CO_3 to add some alkalinity and stabilize the water after chlorination which follows filtration. Offsetting the pH drop from a 6-mg/L Cl_2 dose would cost \$0.02/1000 gal. The total cost of making these changes would be \$0.07/1000 gal and would increase overall production cost by 9.3% and result in a THM reduction of 30% to 40% in the blend and a hardness reduction of 30% from the lime plant and 15% hardness reduction in the blend.

6.1.4 Plant Optimization Summary

Three separate lime-softening plants: AID, VOG, and Venice underwent similar changes to the softening process to reduce THM's. The changes are (1) increasing softening pH to 11.0 to 11.3, (2) adding H_2SO_4 after softening to replace Cl_2 as an acid source for pH reduction and (3) using final chlorination only to maintain a residual. These changes resulted in a 35% to 50% reduction in THM's in the finished water. The corresponding unit cost increase ranged from \$0.02/1000 gal at Venice to \$0.22/1000 gal at VOG

Table 37. Cost Comparisons for the Regular and Optimized Treatment Processed at the Venice Water Treatment Plant[1]

Parameter	Regular Treatment ($/1000 gal)	Optimum Treatment ($/1000 gal)	Increment ($/1000 gal)
Chemical	0.21	0.23	0.02
Labor	0.17	0.17	0.00
Electrical	0.48	0.48	0.00
Total	0.86	0.88	0.02
THM (μg/L)	151	94[2]	

[1] Cost basis for regular Treatment includes regular chemical and labor costs for both the Lime and Reverse Osmosis plant.

[2] Nine-day average of THM's in Venice Distribution system during optimization of lime plant (11/30/84-12/17/84)

Table 38. Cost for the Regular Operation of the Venice Reverse Osmosis and Lime Softening Plants

Parameter	Lime $/1000 gal	Reverse Osmosis $/1000 gal	Total
Chemical	0.23	0.18	0.21
Labor	0.18	0.14	0.17
Electrical	0.34	0.59	0.48
Total	0.75	0.91	0.86
Production (mg/L)	0.94	0.99	1.93
THM (μg/L)	----	----	151

[1] THM's are from the Venice distribution system and represents the blend of finished waters from both plants taken on 11/30/84, immediately prior to lime plant optimization.

which represents a 2.3% to 14% cost increase. Venice, based on this work, has modified their treatment process with FDER approval to feed H_2SO_4 and changed to a higher pH process. They were the only utility to reduce their THM's to regulation. However more utilities would probably try to minimize their THM's if there was a test that operators could utilize to measure THM's and other precursors quickly and inexpensively. It should be noted that during all the plant optimization work, no consumer complaints were received that were directed at the process. Complaints were received by each utility, but these complaints were no more than normal and perhaps even less.

6.2 PACl/ALUM COAGULATION

6.2.1 Introduction

This section of the report describes bench and plant scale work during wet and dry seasons where PACl and Alum were investigated for organic removal. The organic parameters measured were DOC, THM, TOX, and color. The purpose of this work was to report the effectiveness and cost of PACl coagulation for THM reduction in a finished water. As earlier referenced, the plant was the Olga water treatment plant located near Fort Meyers, Florida in Lee County. The raw water utilized was the Caloosahatchee River, a low turbidity high color water formed in a Southwest Florida watershed. The general seasonal variation in raw water quality is shown in Table 39.

6.2.2 Bench Testing

PACl and Alum were tested on a bench scale by jar testing in the wet and dry seasons. The purpose of this testing was to determine the coagulation pH and dose that would maximize organic removal in the plant tests with the intent of minimizing THM formation. The latter required that THM formation curves and chlorine dissipation curves be generated at varying temperatures. Additionally, two PACl coagulants were made available which increased bench scale testing.

The bench scale PACl and Alum work was accomplished in the following order. Dry season samples were first collected and color and DOC isopleths were developed at varying dose and pH for Alum and both PACls. Optimum dose and pH were selected for Alum and PACl, and Cl_2 dissipation and THM formation curves were then generated for the dry season water prepared by optimum Alum and PACl coagulation. The wet season bench scale work only considered two coagulants, Alum and the optimum PACl. However, color, DOC, and THM isopleths were developed for each of these. These isopleths are plots of any desired parameter after jar testing at the pH and coagulant dose for that jar test. When percent removal is plotted corresponding to pH and dose for several jar tests, an array results that allows lines representing a constant percentage of removal to be sketched on the array. Such a sketch is called an isopleth and allows the optimum pH and dose for percent removal to be quickly located on the array. It was not necessary to develop chlorine dissipation curves for the wet season because plant work had demonstrated Cl_2 dose could be minimized on site.

TABLE 39. Seasonal Water Quality for the Caloosahatchee River at Olga, Florida

		Wet Season (June-Sept)	(Oct-May)
pH		6.8-7.3	7.3-8.4
Alkalinity	mg/L as $CaCO_3$	100-110	160-220
Total Hardness	mg/L as $CaCO_3$	100-140	160-240
Calcium Hardness	mg/L as $CaCO_3$	80-120	140-220
Color	cpu	140-240	40- 70
TDS	mg/L	200-300	300-600
DOC	mg/L as C	17-28	12- 17

6.2.3 Dry Season Bench Testing

Because the Olga plant was unique in that it was primarily a lime-softening plant but used an Alum dose of 40 to 50-mg/L in situ with softening at pH 10.3 to 10.6, the first bench work was to evaluate the existing process relative to Alum or PACl coagulation. The color, DOC, and THM's after lime softening in combination with varying Alum or PACl doses as a softening aid is shown in Table 40. At the equivalent dose, the addition of 29-mg/L of Alum reduced color an additional 40% whereas PACl reduced color and 20%. DOC reduction by Alum and PACl was increased approximately 15% as was THM's at the plant pH (10.3 to 10.6). This is surprising because the addition of aluminum based salts was observed to increase organic removal at pHs where aluminum hydrolysis products are thought to be soluble in water treatment.

The results of jar tests conducted using Olga raw water during the dry season are presented in Tables 41 and 42. For the dry season Olga jar test, DOC and color were measured on samples coagulated with varying Alum or PACl doses (29 to 175-mg/L $A12\ (SO_4)_3 \cdot 14\ H_2O$) at varying pH (4.5 to 7.0). From these data the optimum pH for color and DOC reduction was found to be 5.5. An optimum dose was not as apparent although 90% of the DOC removal and virtually 100% of the color removal was realized at a dose of 117-mg/L as Alum. Jar tests for Alum and two forms of PACl (AC 100S and AC 190) were repeated at this pH and using doses up to 292-mg/L as Alum. DOC and color and THM formation analyses were performed on these samples.

Based on these data, Figures 25 through 27 were prepared showing the percent reduction in DOC vs pH and coagulant dose. All three coagulants reduced DOC by more than 60 percent at doses of 175-mg/L as Alum. Overall, Alum reduced DOC slightly better (1 to 3%) than either PACl product. This reduction corresponds to less than 0.5-mg/L DOC. The optimum pH for all coagulants was found to be 5.0-5.5. Although the PACl products appeared to have a broader pH range than Alum for DOC removal, the Alum removals are slightly higher.

Constant percent color reduction curves for the dry season are presented in Figures 28, 29, and 30. These show that all three products were capable of reducing color by 90 percent or more by controlling dose and pH. The optimum pH for color reduction was 5.0 for all of the coagulants which was slightly lower than the DOC optimum. Alum performed slightly better (3-5%) than the PACl products under all conditions and removed color over a broader pH range. Either product could reduce the color to 5 cpu or less.

Results of the dry season jar tests at pH 5.5 are given in Table 42. This table shows that DOC reduction at the lower coagulant doses was slightly greater (5%) using Alum. Coagulant doses greater than 175-mg/L resulted in little if any additional DOC reduction. Color data for these jar tests show the same trend as DOC. Virtually all DOC and color removal was realized at a dose of 117-mg/L Alum. Alum coagulation resulted in lower color at lower coagulant doses while little difference (less than 2 cpu) was noted at higher coagulant doses. Alum and PACl 100S doses over 117-mg/L resulted in little improvement (0-2 cpu) in color reduction. PACl

Table 40. Color, DOC, and THM's Following Lime Softening Jar Tests Using Alum or PACl as Polymer During the Dry Season

Polymer	mg/L $Al_2(SO_4)_3 \cdot 14H_2O$	Reaction pH	Color cpu	DOC (mg/L as C)	THM (μg/L)
None	0	8.1*	71	17.42	986
None	0	9.5	60	17.07	716
None	0	10.0	45	16.02	623
None	0	10.5	30	15.01	515
None	0	11.0	22	11.78	363
None	0	11.5	19	9.23	328
None	0	12.0	8	9.28	326
PACl	29	7.9*	41	14.72	832
PACl	29	9.5	45	14.86	632
PACl	29	10.0	45	14.55	578
PACl	29	10.5	26	12.37	406
PACl	29	11.0	11	10.25	326
PACl	29	11.5	15	8.93	395
PACl	29	12.0	4	7.56	305
Alum	29	7.8*	37	14.36	828
Alum	29	9.5	60	16.59	712
Alum	29	10.0	48	15.63	703
Alum	29	10.5	15	11.81	435
Alum	29	11.0	8	9.56	272
Alum	29	11.5	5	8.61	345
Alum	29	12.0	2	7.52	306

*No Lime Added

Rapid Mix = 1.0 minutes @ 100 rpm
Slow Mix = 20 minutes @ 20 rpm
Settling = 30 minutes
Filtration = 0.45μ A-Glass Fiber
Cl_2 Dose = 20 mg/L for 48 hours

TABLE 41. DOC and Color Remaining by Dose and pH Following Alum, PACl 100s or PACl 190 Coagulation

DOC (mg/L as C)

		pH				
	Dose*	4.5	5.0	5.5	6.0	7.0
Alum	0	19.1	18.4	19.0	19.2	18.7
	29	15.3	14.3	14.8	15.8	16.5
	58	13.8	10.1	9.5	12.5	12.6
	117	9.2	7.3	7.6	7.8	9.9
	175	7.2	6.3	6.5	7.1	8.7
PACl 100S	0	19.1	18.4	19.0	19.2	18.7
	29	15.8	15.2	15.3	16.7	17.3
	58	12.5	11.9	12.1	12.4	14.1
	117	9.9	7.5	7.9	8.8	10.5
	175	8.9	6.5	6.8	7.5	9.5
PACl 190	0	19.1	18.4	19.0	19.2	18.8
	29	14.5	15.2	14.5	16.2	16.7
	58	12.7	12.0	11.1	11.5	13.5
	117	9.8	8.2	8.3	8.2	10.2
	175	9.3	7.1	7.2	7.1	9.1

Color (cpu)

		pH				
	Dose*	4.5	5.0	5.5	6.0	7.0
Alum	0	46	82	68	68	71
	29	46	27	38	46	46
	58	44	5	16	24	24
	117	39	2	9	13	13
	175	3	2	9	9	9
PACl 100S	0	46	82	68	68	71
	29	20	16	38	60	53
	58	14	9	20	24	46
	117	23	5	16	13	20
	175	10	5	9	5	13
PACl 190	0	46	82	668	68	71
	29	47	38	31	46	49
	58	38	27	24	20	24
	117	28	9	16	13	20
	175	16	5	13	9	13

*All coagulant doses are as mg/L $Al_2(SO_4)_3 .14H_2O$

TABLE 42. Color, DOC, THMFP and Al residual from PACl/Alum Jar Tests during the Dry Season

Olga Dry Season Jar Tests at pH 5.5

	Dose*	DOC mg/L	Color CPU	THMFP ug/L	Al mg/L
Alum	0	18.5	52	980	0.21
	29	13.0	18	690	0.43
	58	9.55	8	460	0.48
	117	7.32	5	460	0.38
	175	6.74	4	410	0.18
	252	5.73	4	400	0.19
PACl 100S	29	14.3	28	950	0.39
	58	11.4	14	680	0.36
	117	8.09	6	570	0.32
	175	7.04	5	470	0.16
	292	5.95	4	410	0.26
PACl 190	29	14.3	26	950	0.42
	58	11.1	15	870	0.49
	117	8.58	10	620	1.35
	175	7.12	8	540	2.23
	292	6.24	6	440	2.13

*All coagulant doses are as mg/L $Al_2(SO_4)_3 \cdot 14H_2O$

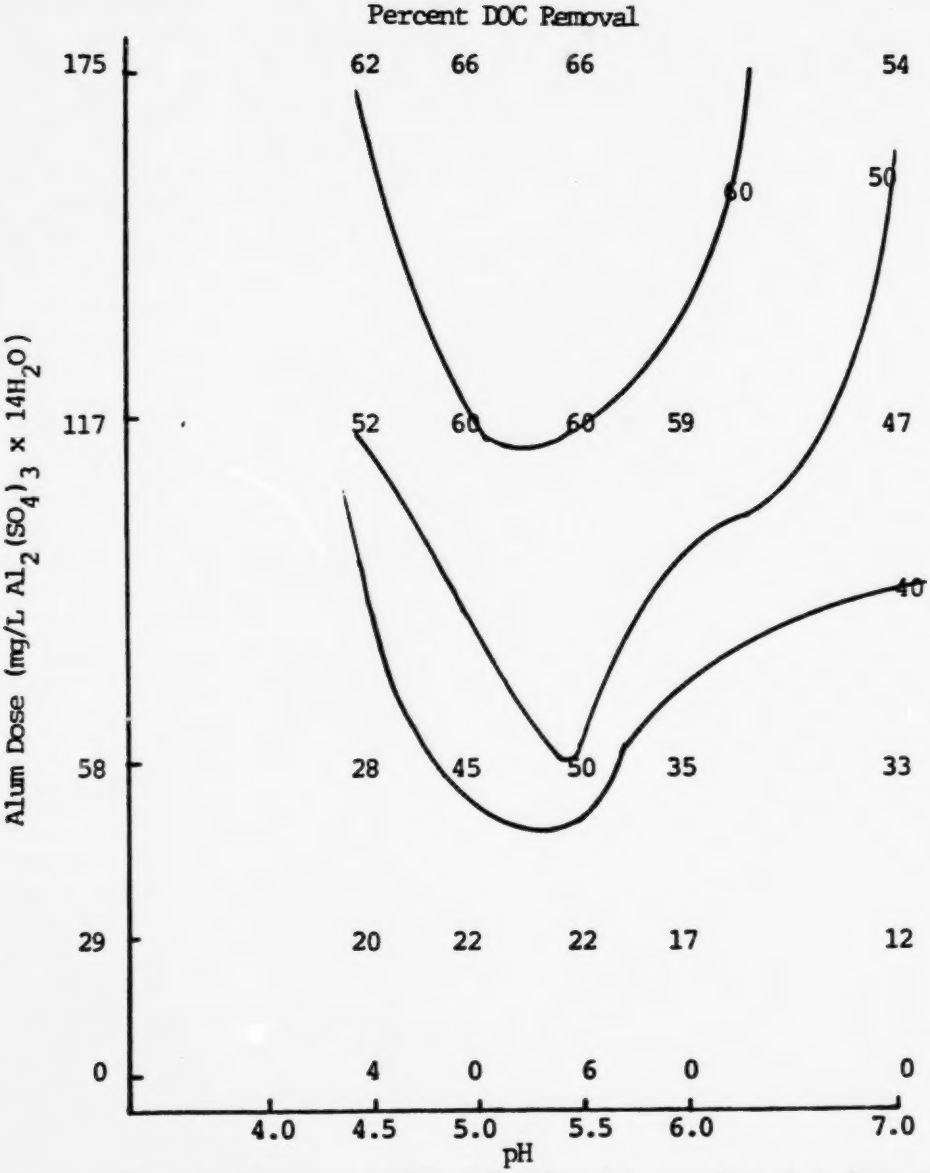


Figure 25 Dry season percent DOC reduction for varying alum dose and pH using Olga Raw Water (DOC = 19 mg/L).

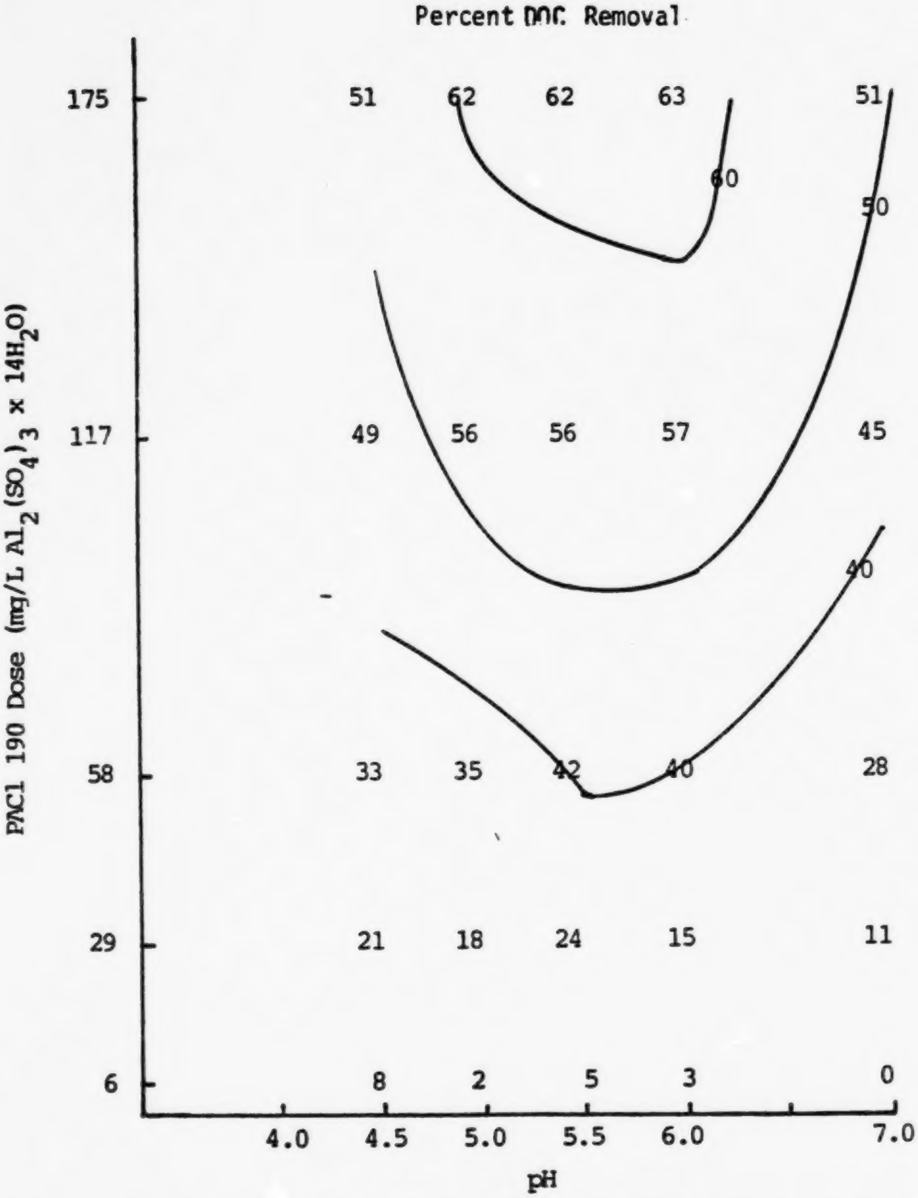


Figure 26 Dry season constant percent DOC reduction curves for varying PACl 190 dose and pH using Olga raw water. Raw water DOC = 19.0 mg/L.

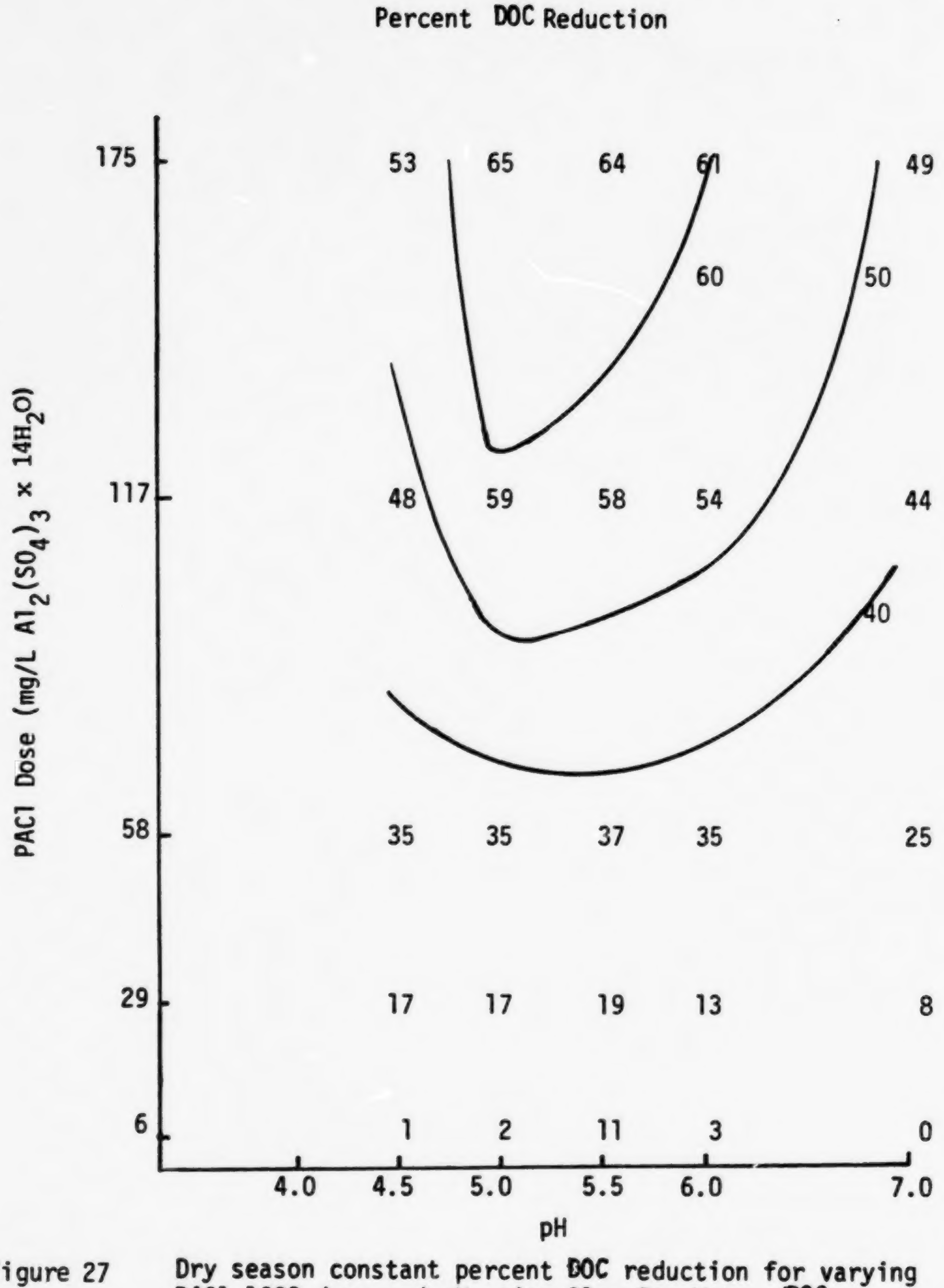


Figure 27 Dry season constant percent DOC reduction for varying PACl 100S dose and pH using Olga Raw Water (DOC = 19 mg/l)

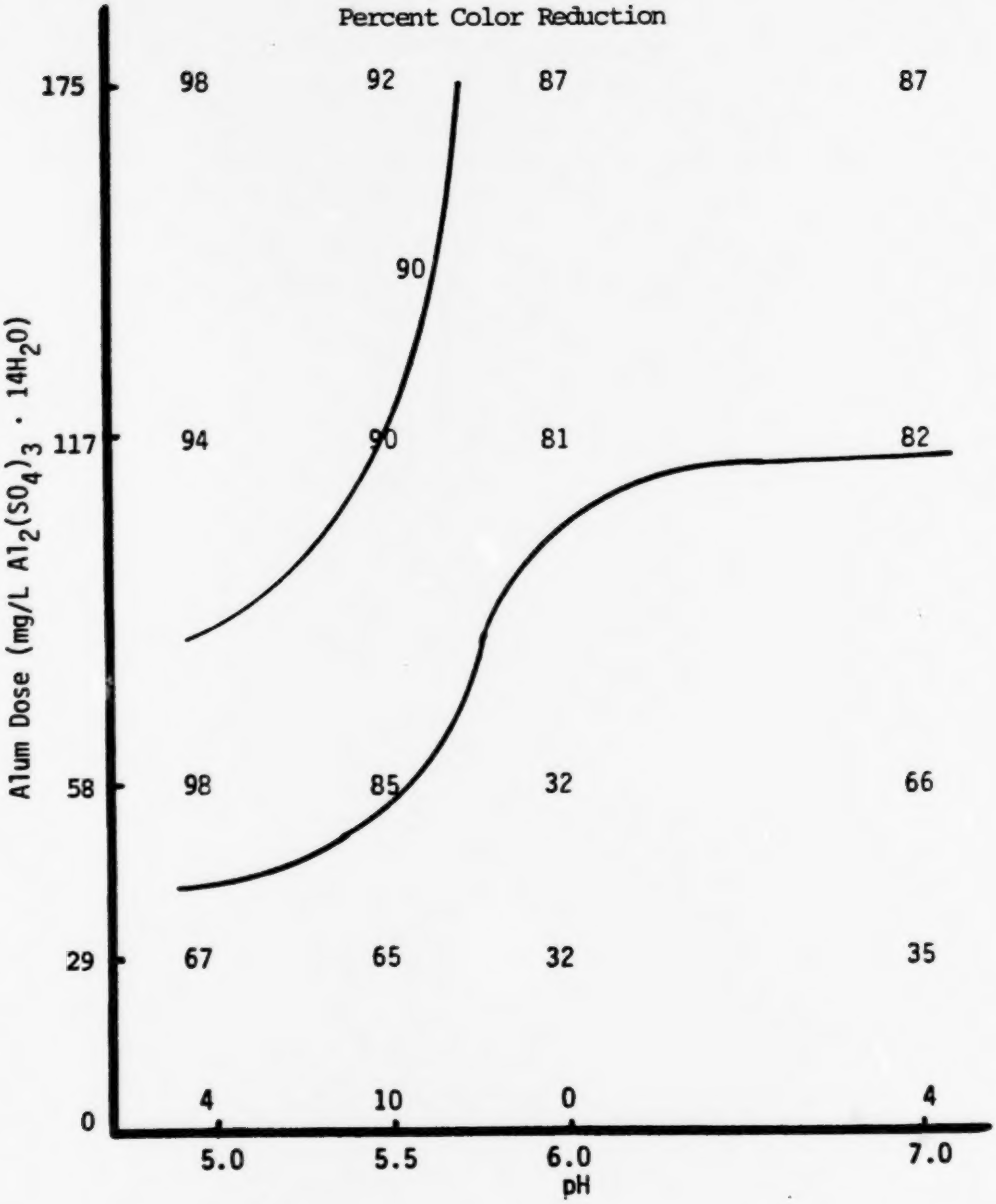


Figure 28 Dry season constant percent color reduction curves for varying alum dose and pH (Initial color=70 cpu).

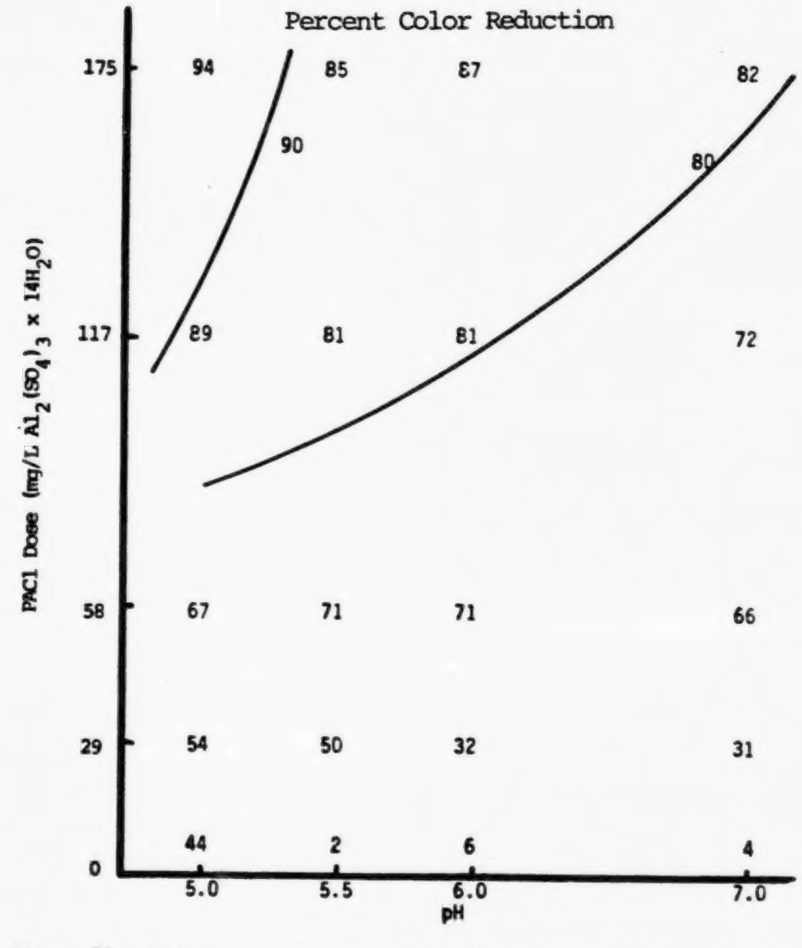


Figure 29 Dry season percent color reduction curves for varying PACl-190 dose and pH (Initial Color=70cpu)

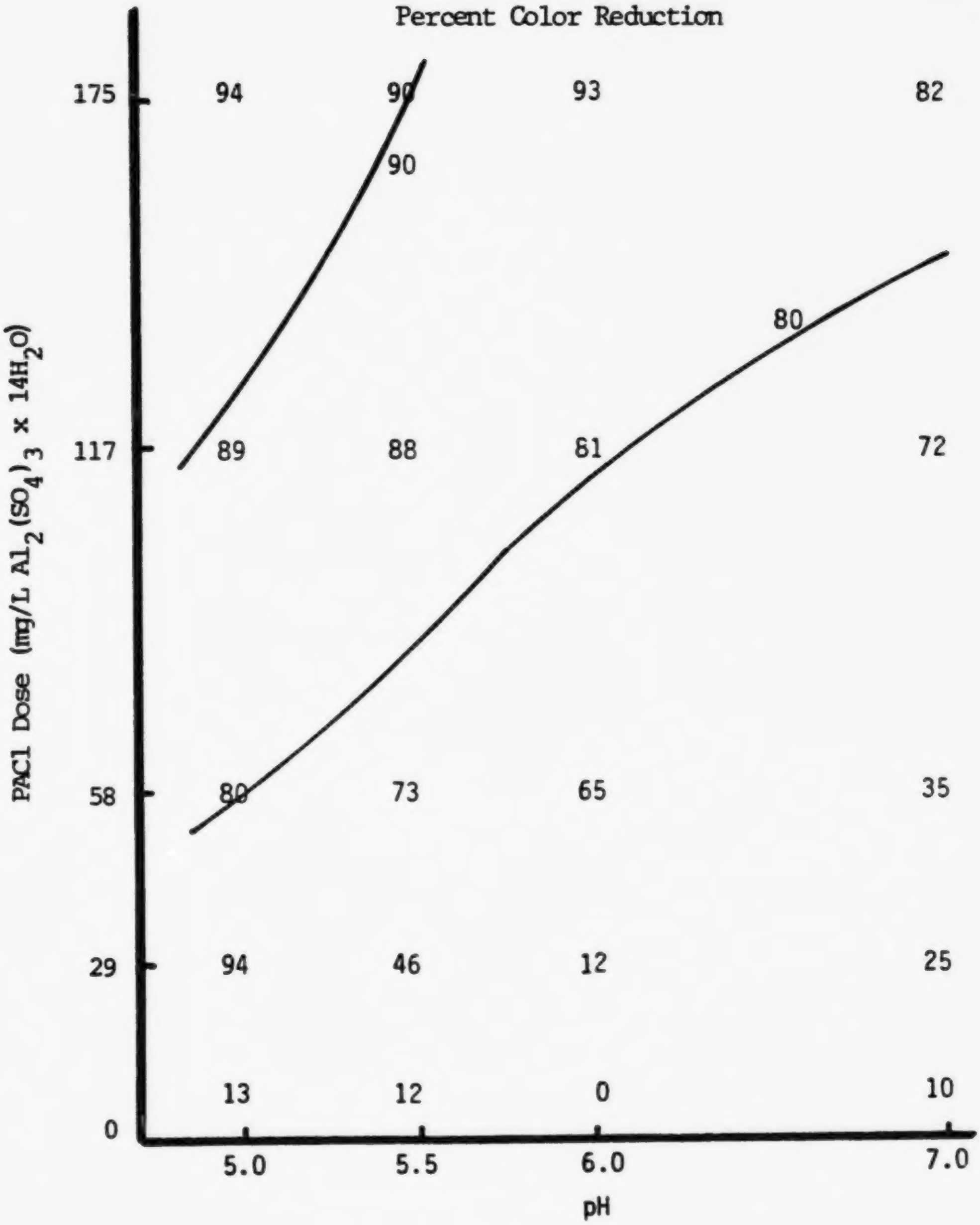


Figure 30 Dry season constant percent color reduction curves for varying PACl-100S dose and pH (Initial Color=70 cpu).

190 color reduction leveled off at 117-mg/L as Alum. All of the coagulants could reduce color to less than 15 cpu using doses of 58-mg/L as alum or more.

THM formation was reduced by almost 50 percent using alum doses as low as 58-mg/L, although little further reduction was observed as the dose increased. THM's reduction was not as great for either of the PACl products save the highest coagulant dose (292-mg/L).

Aluminum residuals at pH 5.5 increased slightly over background levels at low alum and PACl 100S doses and decreased to background levels at a dose of 292-mg/L. Typically Al residuals were less than 0.4-mg/L as Al. PACl 190 aluminum residuals however, continued to increase with increasing coagulant dose. The aluminum residuals in the PACl 190 samples represented as much as 10 percent of the added aluminum. The Al residuals indicate that the organics react with the Al species and are not removed until the Al exceeds the organic demand and is available to form a settleable floc. Once the optimum pH and dose had been established for the dry season jar tests, THM formation curves as shown in Figures 31, 32, and 33 were generated for Cl_2 doses dose of 8, 16, 32-mg/L. The THM's generated from a Cl_2 dose of 8-mg/L appears to be approaching a limit around 400 µg/L at 120 hours. THM's corresponding to Cl_2 doses of 16 and 32-mg/L do not approach stability after 120 hours. This data indicates that THM formation would be minimized by using alum or PACl 100S and the lowest Cl_2 doses possible to maintain a 0.2-mg/L Cl_2 residual.

The chlorine dissipation data indicates that 8-mg/L Cl_2 was adequate to maintain a residual after 120 hours. Because of additional chlorine demands in plant storage and the distribution system a Cl_2 dose of 8 to 16-mg/L was selected for plant optimization and PACl testing.

The data from the THM's and Cl_2 dissipation curves for the Alum and PACl prepared waters were contrasted against plant samples for chlorine demand, color and THM formation. A sample was collected from the plant immediately after chlorination (20-mg/L) and the resulting Cl_2 dissipation, THM formation and color are presented in Figures 34, 35, and 36. This data shows the chlorine demand and the THM formation were much higher for the plant samples compared to the jar samples. A Cl_2 residual of 1-mg/L was available at 120 hours for the plant samples following an initial dose of 20-mg/L whereas this residual was obtained in jar samples for a dose of 8-mg/L Cl_2. The THM formation was in excess of 800 µg/L for the plant samples and did not exceed 600 for jar samples. Color dissipated as expected with time in the plant sample due to the Cl_2 residual. This data relative to the bench scale work indicated a major reduction in THM formation should occur during the plant study.

The dry season PACl and alum jar tests showed that pH 5.5 should optimize organic removal. Color, DOC, and THM precursor removals were closely related with percent DOC and THM precursor removal being about equal. The general order of coagulants based on highest organic removal was alum, PACl 100S and PACl 190 although the difference between alum and

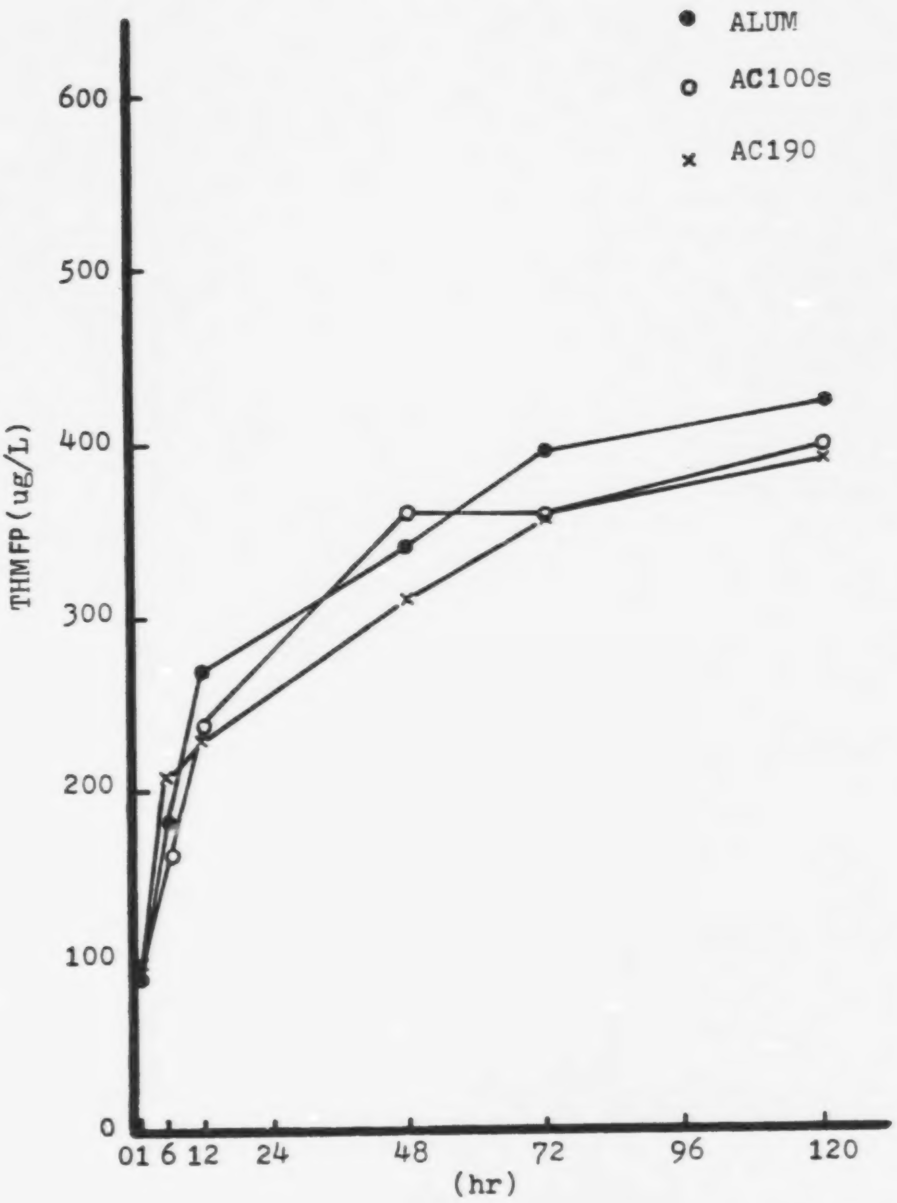


Figure 31 THMFP vs Time for Olga raw water treated in the laboratory with 175 mg/L coagulant at pH = 5.5. Cl_2 = 8 mg/L

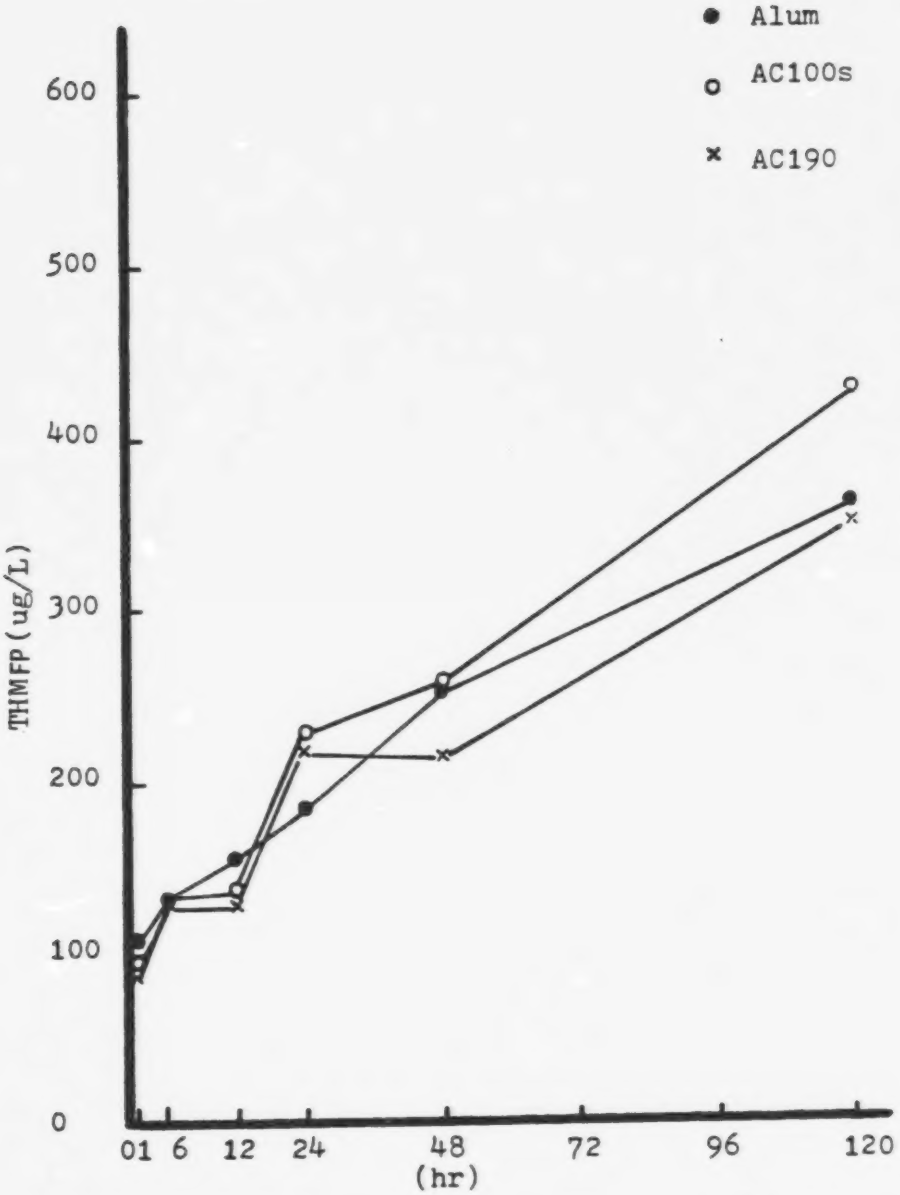


Figure 32 THMFP vs. Time for Olga raw water treated in the laboratory with 175 mg/L coagulant at pH = 5.5. Cl_2 = 16 mg/L.

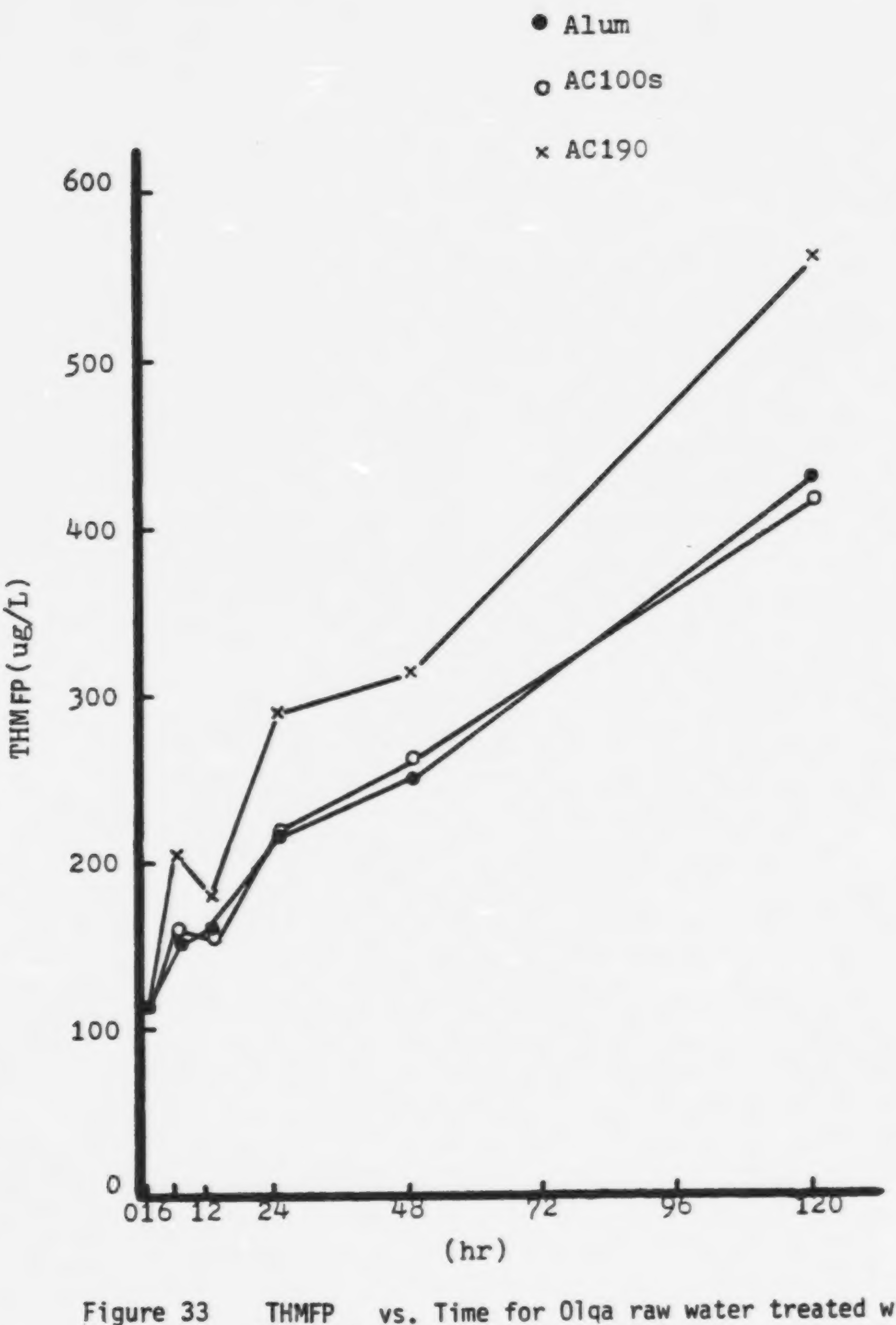


Figure 33 THMFP vs. Time for Olqa raw water treated with 175 mg/L coagulant at pH = 5.5. Cl_2 = 32 mg/L.

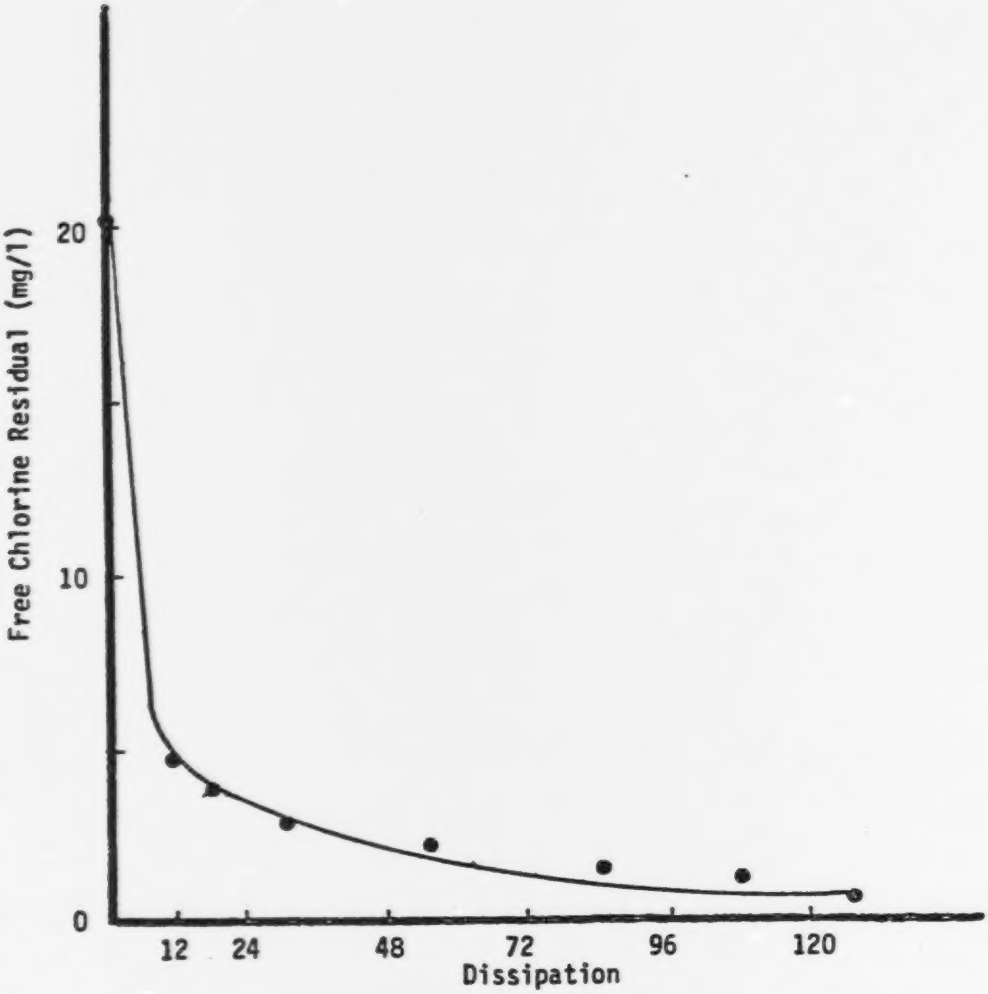


Figure 34   Chlorine Residual Olga Treated Water 25°C.

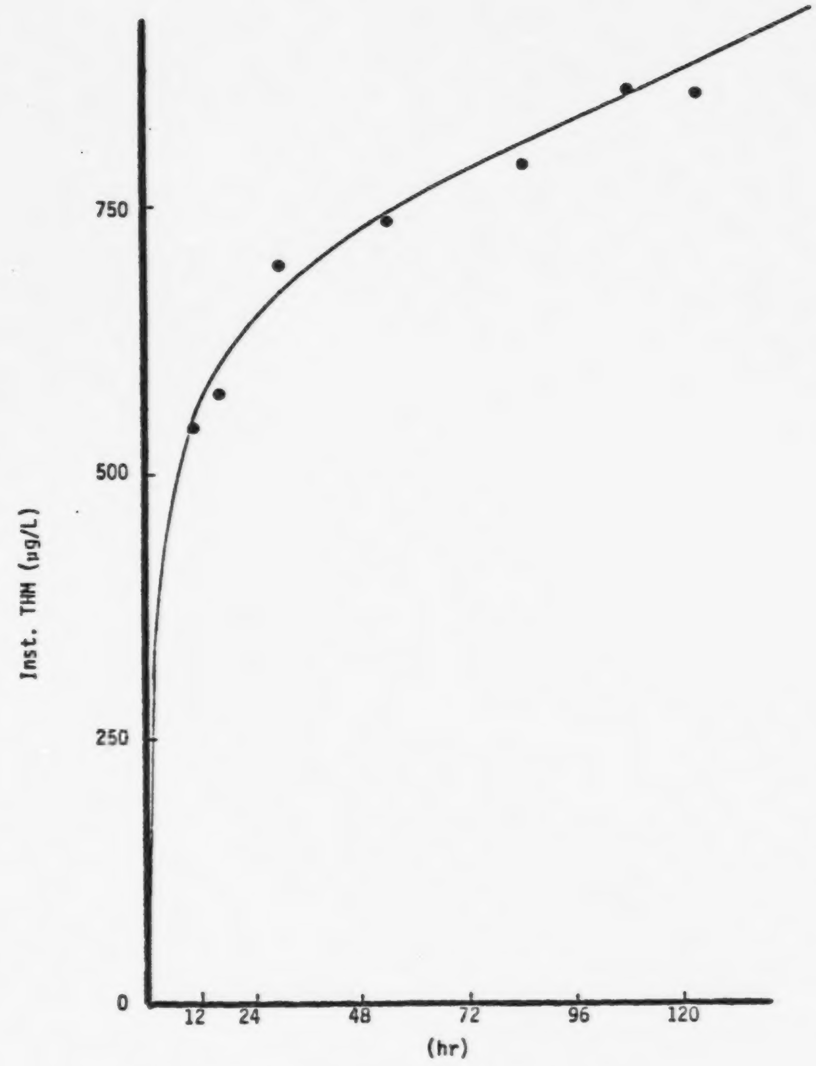


Figure 35 Trihalomethane Formation vs time.
Olga Treated Water 25°C.

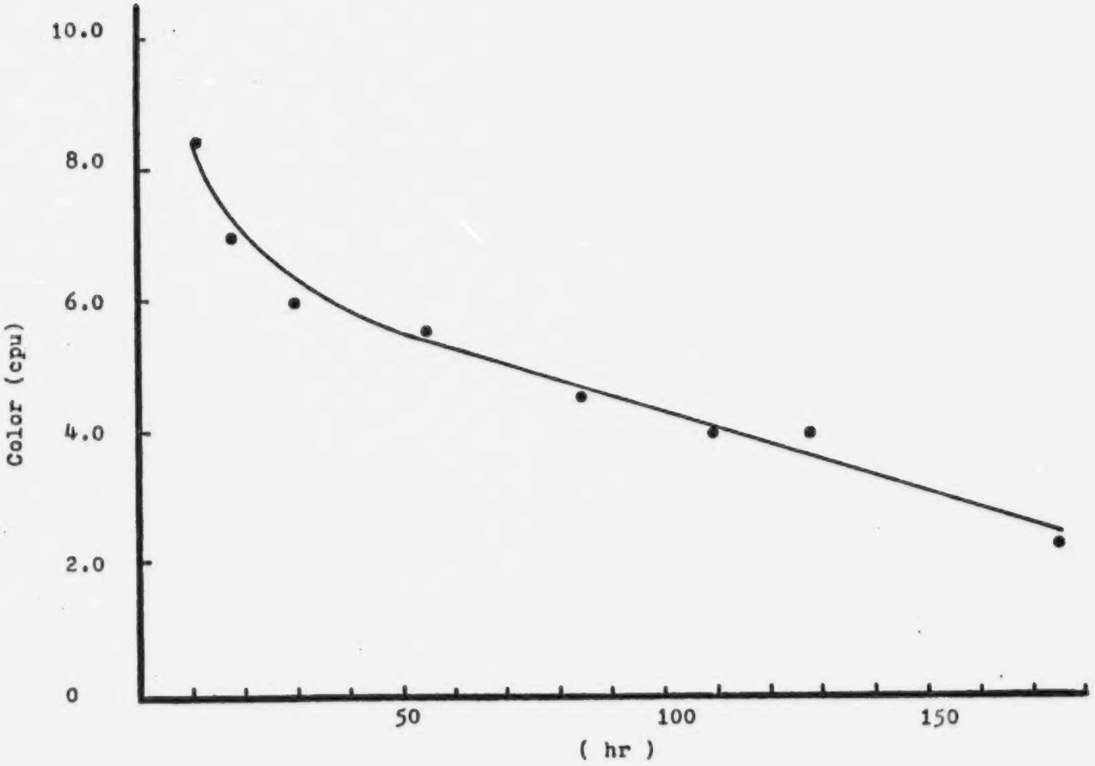


Figure 36 Residual Color vs. Time for Olga Finished Water

PACl 100S is very slight. Based on these results the latter two coagulants were selected for plant testing.

6.2.4 Base Neutralizing Capacity

The base neutralizing capacity of alum, PACl 100 and PACl 190 were compared to H_2SO_4 by a standard alkalinity titiation. The results of this work are presented as standard titration curves for a common water in Figure 37. The optimum pH for organic removal was found to be 5.5 during the dry season jar testing. This is not uncommon as many alum coagulation plants have found the pH of maximum Al insolubility correspond to maximum color removal during coagulation. This is very convenient for the plant operator as he doses alum, destroying alkalinity until a coagulation pH of 5.5 to 7.0 is reached and generally achieves very good color removal. The alkalinity of the Olga raw water was found to be 90-mg/L at pH 4.5, the HCO_3 endpoint. The optimum pH 5.5 was reached after 1.4 meq/L acid had been added when H_2SO_4 was used.

The meq/L of Alum, PACl 100S and PACl 100 required to reach pH 5.5 were 1.5, 3.0, and 3.3. This means that if a large alum dose were required to obtain maximum organic removal and little alkalinity was available, PACl would possibly be advantageous over Alum. Alum would require additional alkalinity whereas PACl would not. This situation is reversed where raw alkalinity is high relative to initial dose. Obviously a base neutralization capacity is a factor that must be considered when coagulation pH and dose are important.

6.2.5 Wet Season Bench Testing

Results from the wet season jar tests are given in Tables 43 and 44. These tests were performed after and utilized the results of the dry season jar tests. The wet season analyses include DOC, THM's, and color isopleths as well as Al residuals. These analyses were repeated plus TOX at the optimum pH for THM's removal.

Constant percent DOC removal curves are shown in Figures 38 and 39 for alum and PACl. Optimum DOC reduction for both coagulants occurred at pH 5.0-5.5 with over 70 percent of the raw DOC removed at a dose of 175-mg/L. Little additional DOC reduction was achieved by either coagulant by increasing the dose to 292-mg/L. The effectiveness of both coagulants for removing DOC decreased as pH was increased above 6.0, particularly at the higher coagulant doses. The lowest DOC value found during the jar testing for alum was 4.42-mg/L. This occurred when alum was used at 292-mg/L and at pH 5.5 and removed 79% of the initial DOC. The lowest DOC produced by PACl treatment occurred with a dose of 292-mg/L (as alum) and pH 5.0, resulting in a DOC of 5.21-mg/L which was 76% DOC removal.

The effect of coagulant dose and pH on THM's reduction was similar to the dose/pH effects of DOC reduction (see Figures 40 and 41). A dose of 175-mg/L at pH 5.0 to 6.0 removed 60 to 70% of the THM's and represented optimum treatment for THM's as well as DOC. At the highest coagulant doses both PACl and alum resulted in approximately 200 µg/L THM formation when

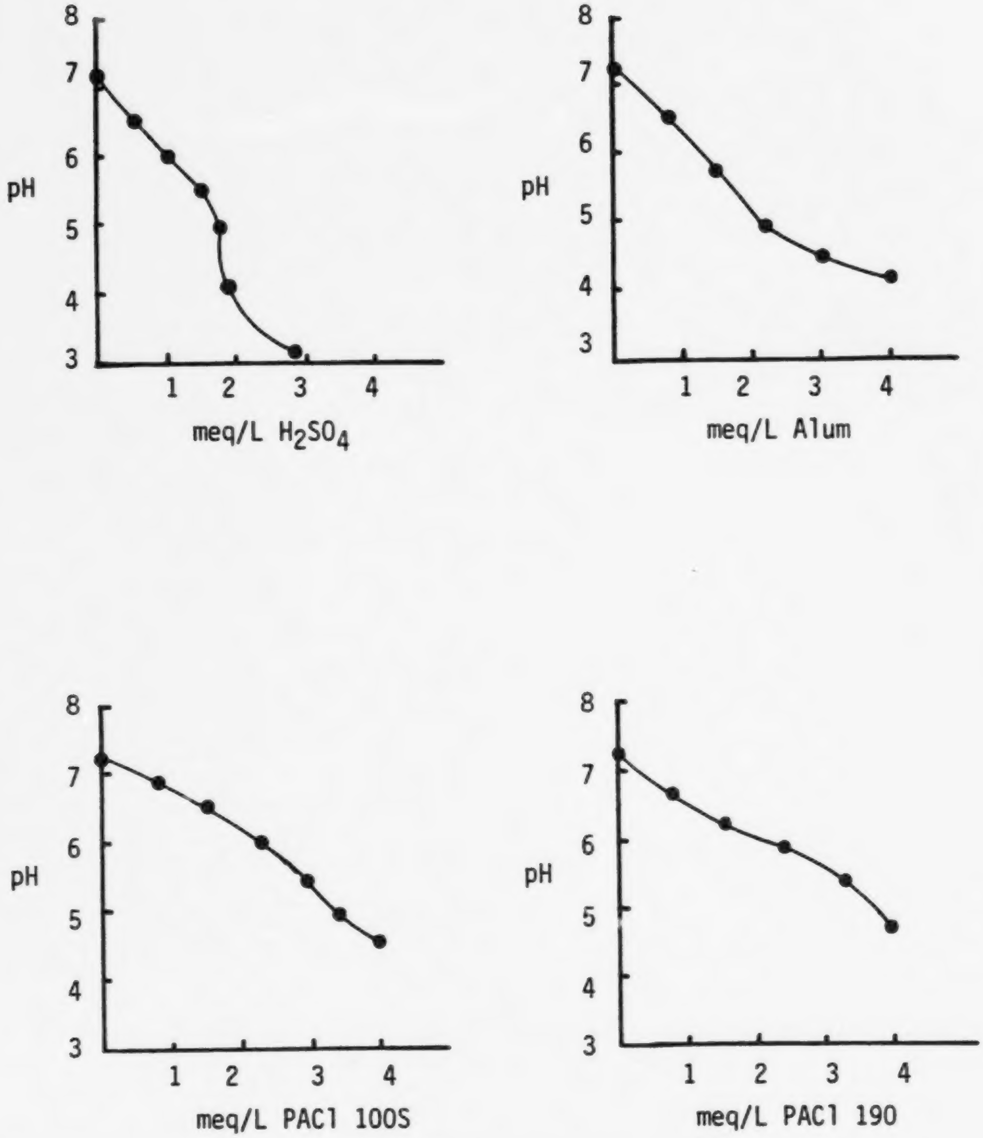


Figure 37 Titration curves demonstrating base neutralization capacity of Alum, PACl 100S and PACl 190 relative to H_2SO_4.

Table 43. DOC, THMFP, TOXFP, Color and AL residuals Following Alum Coagulation Jar Tests During the Wet Season

pH	Alum Dose (mg/L $Al_2(SO_4)_3 \cdot 14H_2O$)	DOC (mg/L as C)	THM's (μg/L as Cl)	TOX (μg/L)	Color (cpu)	Al (mg/L)
	CONTROL	22.28	595	1988	100	0.10
4.5	29	12.51	585		43	1.15
	58	12.50	495		34	2.13
	117	7.92	320		24	0.58
	175	6.60	275		15	2.00
	292	5.01	160		15	5.15
5.0	29	15.92	340		43	0.53
	58	10.78	285		29	0.24
	117	7.62	315		21	0.13
	175	6.28	230		9	0.18
	292	4.78	260		7	0.23
5.5	29	16.62	515	1832	51	0.41
	58	11.21	480	1339	24	0.18
	117	7.81	430	922	11	0.15
	175	5.99	190	760	6	0.15
	292	4.42	195	699	5	0.13
5.5 Duplicate	29	16.37	460	1845	49	0.44
	58	11.04	310	1306	24	0.24
	117	8.25	245	966	10	0.21
	175	6.10	300	----	6	0.14
	292	4.85	140	670	5	0.13
6.0	29	18.29	485		64	0.90
	58	11.98	380		25	0.24
	117	8.97	295		9	0.17
	175	6.80	225		7	0.13
	292	6.64	180		7	0.13
6.5	29	18.00	395		69	0.80
	58	12.40	400		29	0.20
	117	9.49	335		12	0.15
	175	7.82	315		9	0.15
	292	7.30	250		9	0.14
7.0	29	19.20	540		80	0.22
	58	14.12	490		28	0.15
	117	10.66	335		16	0.14
	175	8.96	305		12	0.19
	292	8.38	330		10	0.14
8.0	29	20.00	575		65	0.37
	58	16.47	630		45	0.35
	117	13.61	470		25	0.33
	175	10.23	500		23	0.31
	292	10.99	430		20	0.34

Table 44. DOC, THMFP, TOXFP, Color and AL residuals Following PACl 100S Coagulation Jar Tests During the Wet Season

pH	Alum Dose (mg $Al_2(SO_4)_3$)	DOC (mg/L as C)	THMFP (μg/L as Cl)	TOX (μg/L)	Color (cpu)	Al (mg/L)
	CONTROL	21.80	920		90	0.19
4.5	29	17.10	645		52	0.68
	58	13.16	575		31	0.84
	117	8.80	495		14	1.33
	175	7.90	285		13	0.18
	292	5.58	335		12	2.19
5.0	29	15.24	555		47	0.58
	58	12.45	570		25	0.33
	117	8.00	405		14	0.18
	175	6.60	250		8	0.13
	292	5.21	255		8	0.12
5.5	29	15.11	875	----	45	0.49
	58	13.37	915	1790	30	0.19
	117	8.86	375	1205	9	0.16
	175	7.30	345	951	6	0.12
	292	5.44	185	859	3	0.13
5.5	29	15.80	775	2423	58	0.57
Duplicate	58	13.34	355	2027	30	0.29
	117	8.98	305	1466	9	0.13
	175	6.88	290	1046	6	0.13
	292	5.38	205	940	3	0.12
6.0	29	15.89	725		44	0.52
	58	15.34	720		40	0.34
	117	10.07	390		10	0.12
	175	7.55	315		6	0.12
	292	6.61	325		4	0.12
6.5	29	18.62	740		75	0.93
	58	15.34	675		40	0.19
	117	11.41	450		15	0.14
	175	10.07	440		11	0.12
	292	7.40	365		6	0.11
7.0	29	16.39	610		60	0.20
	58	16.03	700		47	0.15
	117	12.82	725		20	0.15
	175	10.16	430		11	0.13
	292	8.17	330		8	0.13
8.0	29	19.50	795		68	0.34
	58	18.68	760		55	0.38
	117	15.46	635		27	0.31
	1/5	13.69	545		27	0.30
	292	17.57	595		19	0.34

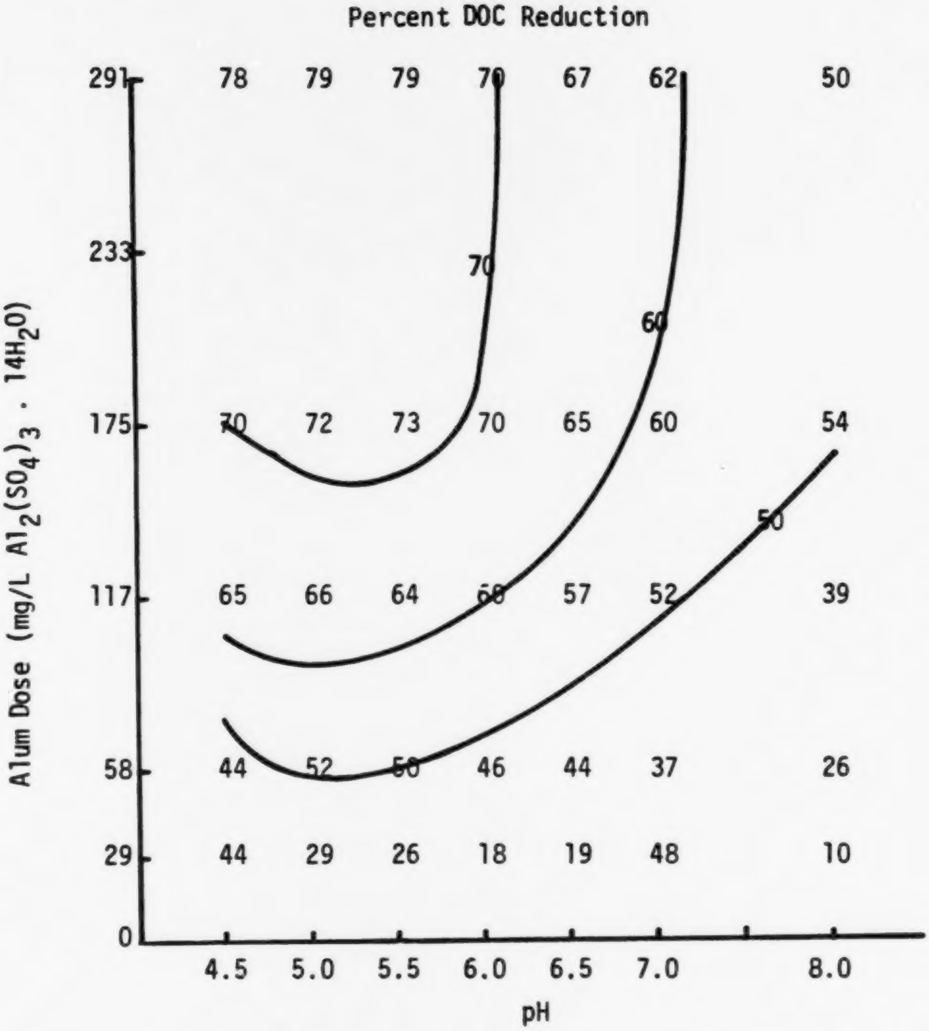


Figure 38 DOC Isopleths for Alum Coagulation during the Wet Season (Raw DOC 22.3 mg/L).

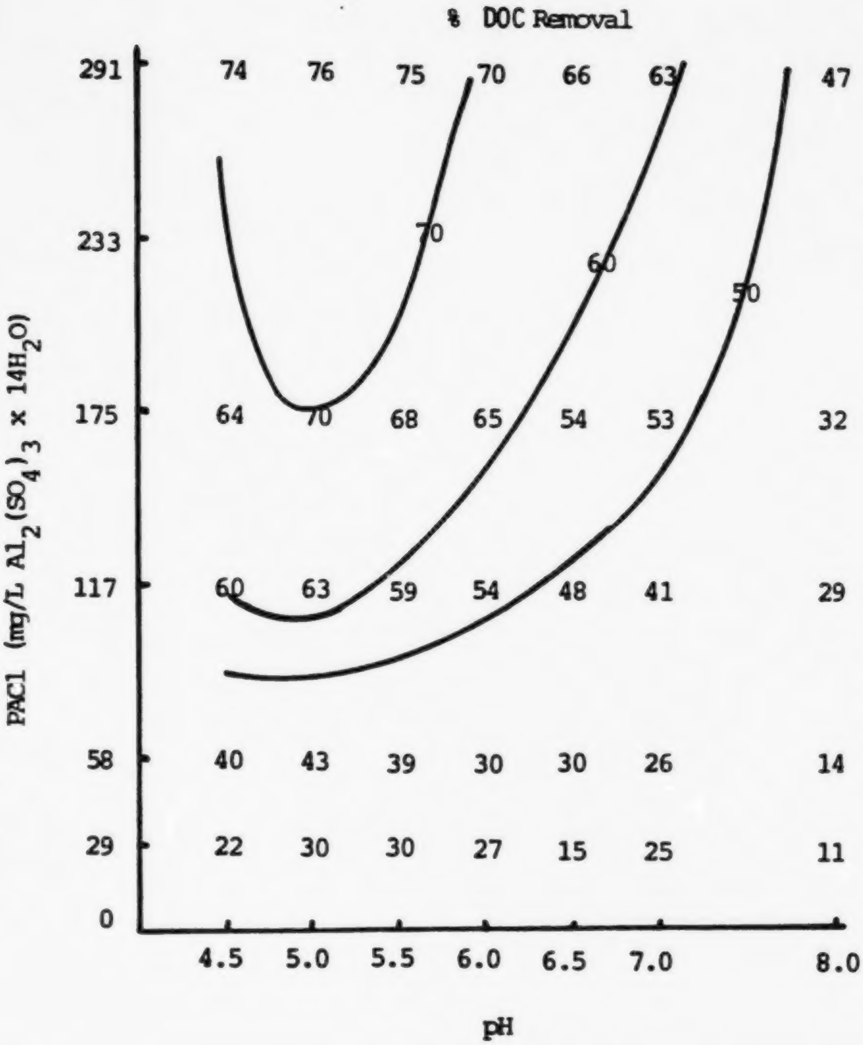


Figure 39 DOC Isopleths for PACl-100S Coagulation during wet season (Raw DOC=22.3 mg/L).

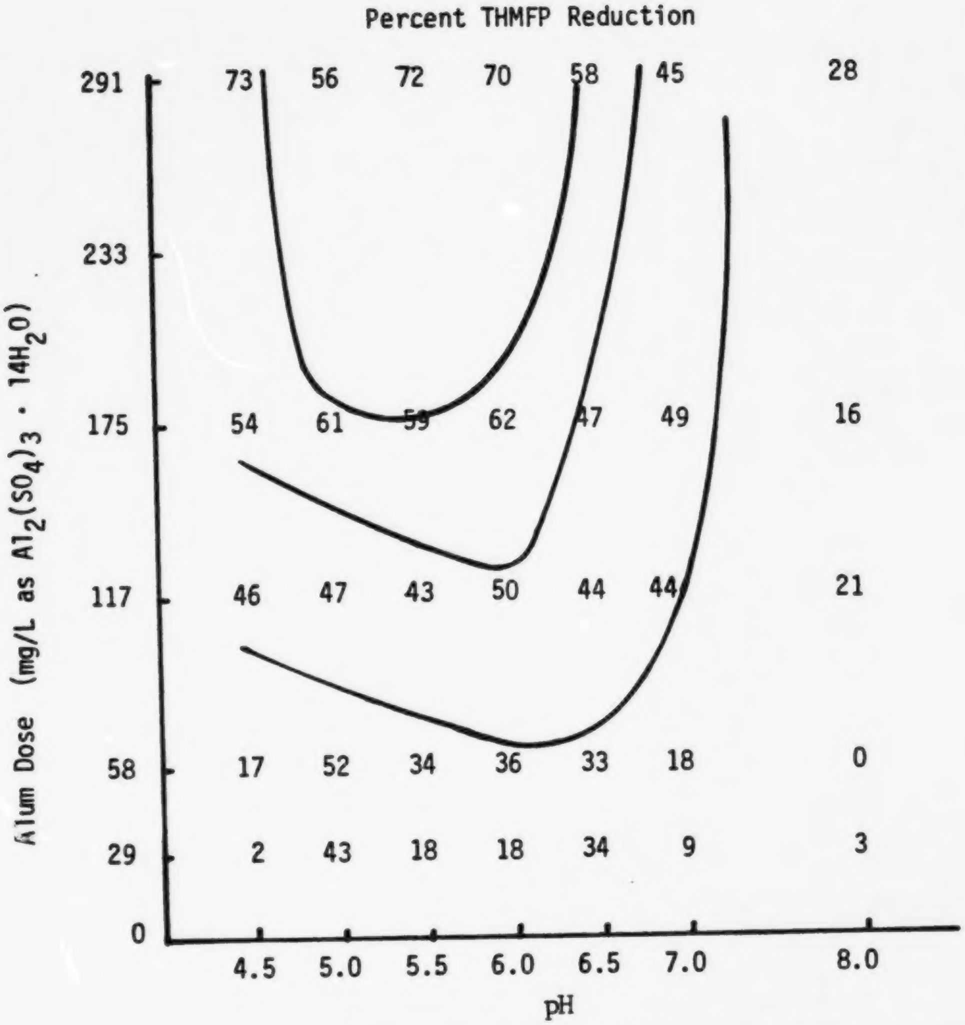


Figure 40 THMFP isopleths for Alum coagulation during wet season (Raw THMFP = 595 ug/L)

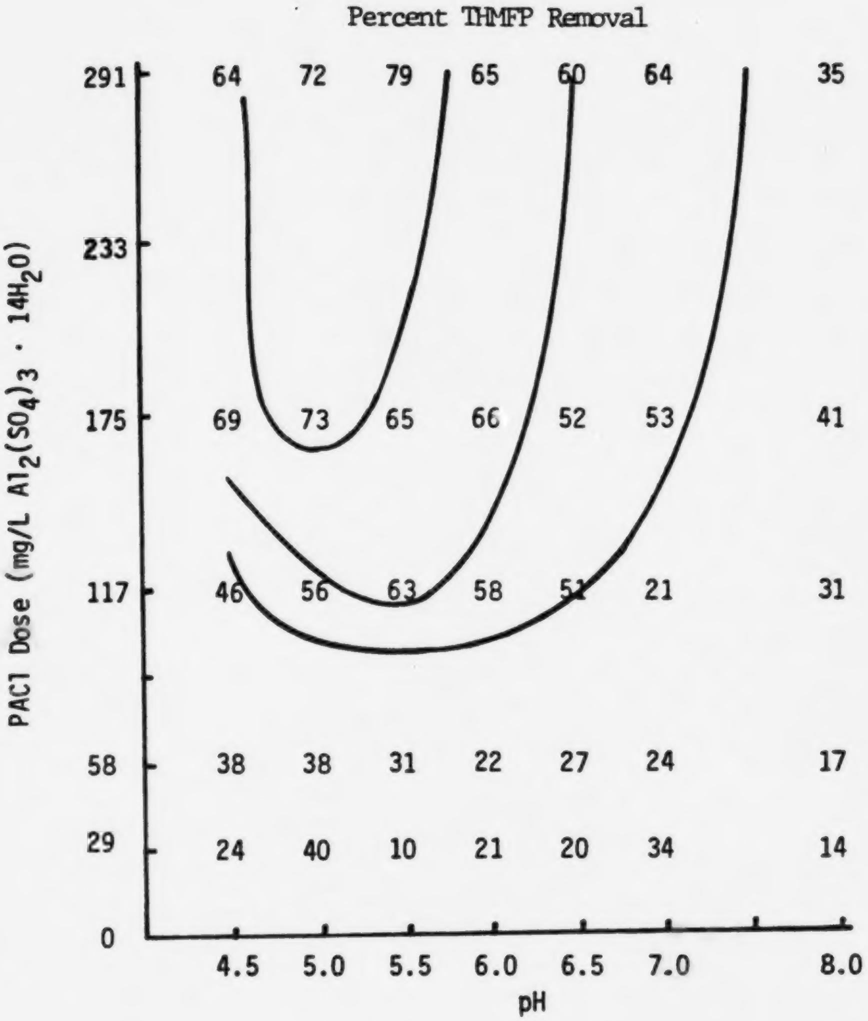


Figure 41 Isopleth for percent THMFP removal after PACl 100S coagulation during wet season. Raw THMFP = 920 ug/L.

chlorinated for 48 hrs at 20°C using 20-mg/L chlorine at pH 7.5. PACl resulted in higher percent removals but also higher residuals, since the raw water used in those jar tests had a higher formation potential than that used with alum coagulation. Both coagulants showed optimum performance at pH 5.5 with relatively little improvement between doses of 175 and 292-mg/L. In both cases performance deteriorated rapidly as pH increased above 6.0. TOX formation was measured for samples coagulated at pH 5.5 for both alum and PACl. These results are given in Figure 42 and range from 2000 μg/L to 700 μg/L. Coagulant doses greater than 175-mg/L had little effect on TOX formation, however alum coagulation resulted generally in lower levels (10 to 15%) of TOX than PACl coagulation. Constant percent color reduction curves for the wet season jar tests are shown in Figures 43 and 44. Both coagulants reduced color to less than 15 cpu with doses of 117-mg/L as alum. The optimum pH for color reduction was between 5.5 and 6.0 for both coagulants. The constant percent reduction isopleths were broader for color than for DOC or THM formation and indicated that color causing organics are more easily removed by coagulation than some colorless DOC forms or THM precursors.

Aluminum residuals in the wet season jar tests generally increased up to a dose of 29 to 58-mg/L and decreased to background levels at higher doses, except for those jars at pH 4.5 or 8.0 The pH 4.5 tests resulted in aluminum residuals as high as 5.15-mg/L while the pH 8.0 tests never exceeded 0.4-mg/L. Even at optimum pH for aluminum precipitation, 5.5, elevated aluminum residuals were found in the product water. The high aluminum residuals coupled with low DOC reductions at low coagulant doses suggest that aluminum is complexed by the organic compounds present in the raw water. Aluminum residuals in PACl or Alum coagulated waters in a normal pH range of 5.5 to 7.5 should not exceed 0.5-mg/L and should not present any acute problem concerning Al residuals to consumers.

Figures 45 and 46 are three-dimensional plots of residual Al, coagulation pH and dose for alum or PACl 100S. This figure was created to demonstrate the mechanism of organic removal by coagulation with aluminum salts. As is seen in Figure 45, the lower dose of 29-mg/L alum has a higher Al residual than higher alum doses at a constant coagulation pH. The highest residual is 2.2-mg/L Al at pH 4.5. For a constant dose the Al residual decreases at pH 5.0 and 5.5 before increasing at pH 6.0 and 6.5. This covers the range of optimum organic removal. Very low Al residuals are obtained at all coagulation pHs (save 4.5) as the alum dose is increased. The data presented in Figures 45 and 46 indicate a chemical reaction between Aluminum hydrolysis products and soluble organics followed by organic removal by sweep coagulation. The removal of these complexed organics results from the formation of settleable precipitate which is probably mostly $Al(OH)_3$. Additional organic removal occurs as the dose is increased in this pH range probably by adsorption onto the $Al(OH)_3$ floc.

The dry season jar tests had indicated that organic removal was essentially maximized at pH 5.0 to 6.0 and doses of 175-mg/L for DOC, THM's or color reduction. DOC, THM's and TOX removals were very similar at equivalent dose and pH conditions. Color removal approached 90% whereas DOC and THM's could generally be reduced by 70%. Alum was slightly superior to

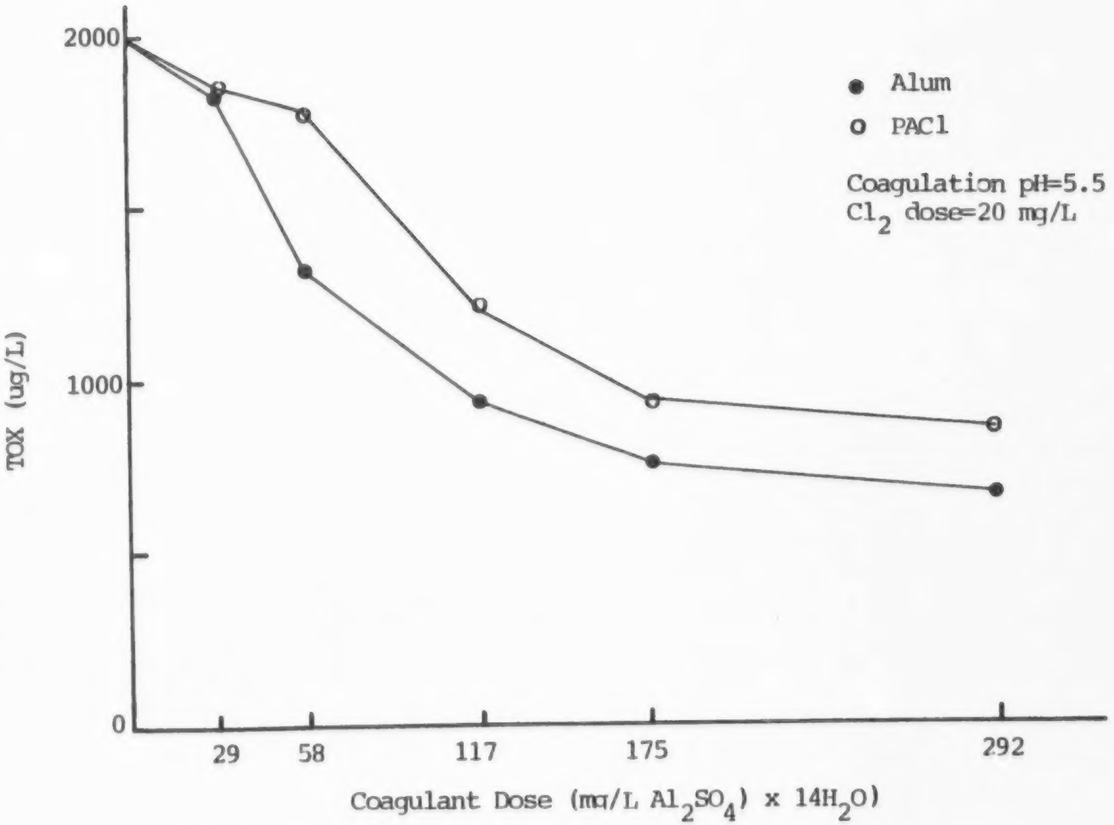


Figure 42 TOX formation, Olga plant wet season jar tests.

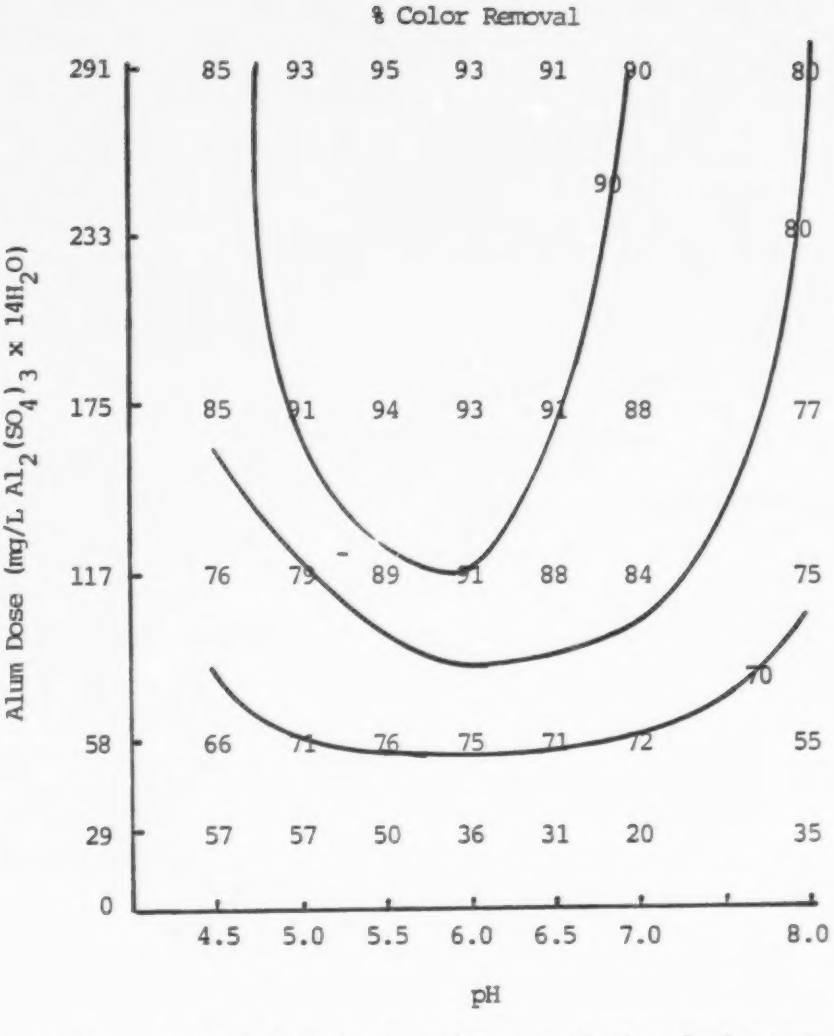


Figure 43 Color Isopleths for Alum coagulation during wet season (Color=100 cpu).

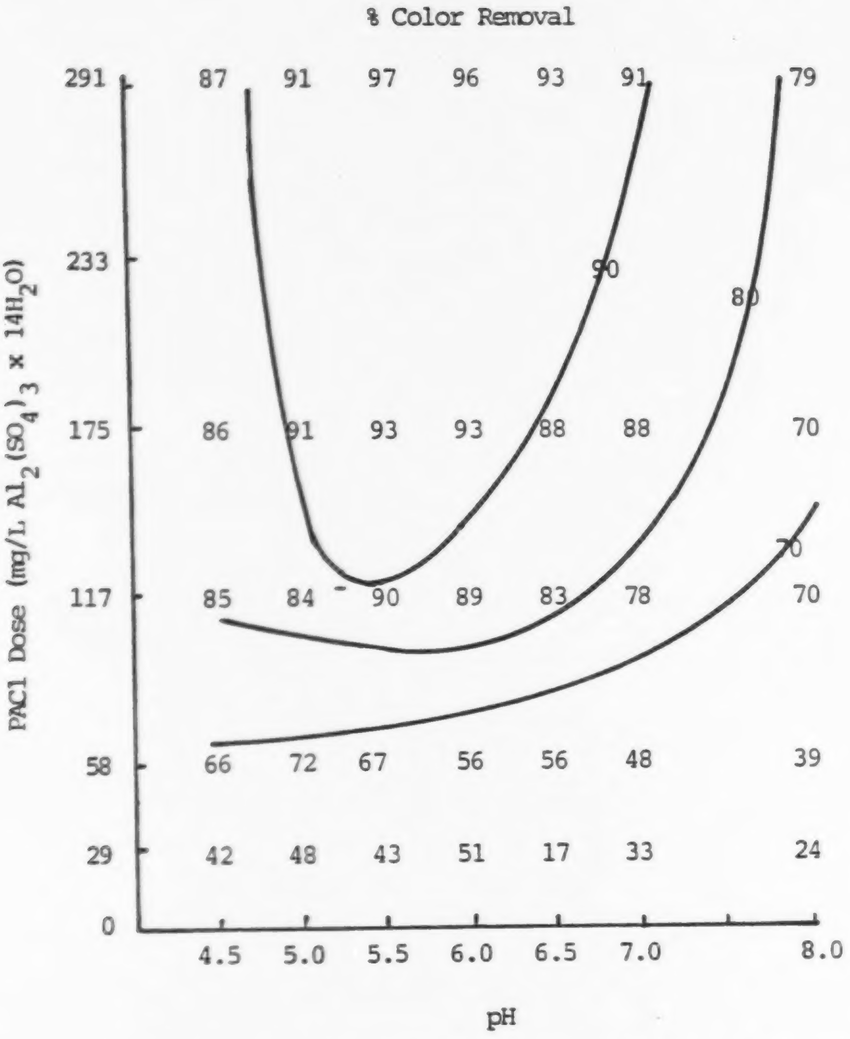


Figure 44 Color Isopleths for PAC1-100S coagulation during wet season (Color=100 cpu).

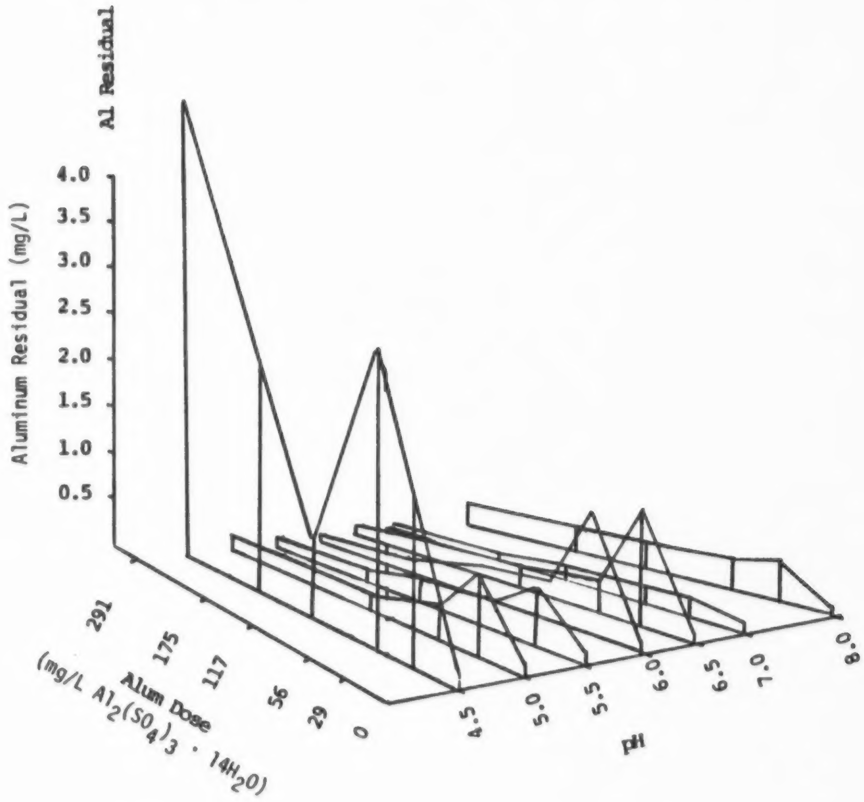


Figure 45 Oiga wet season jar tests aluminum residual vs. alum dose and coagulation pH.

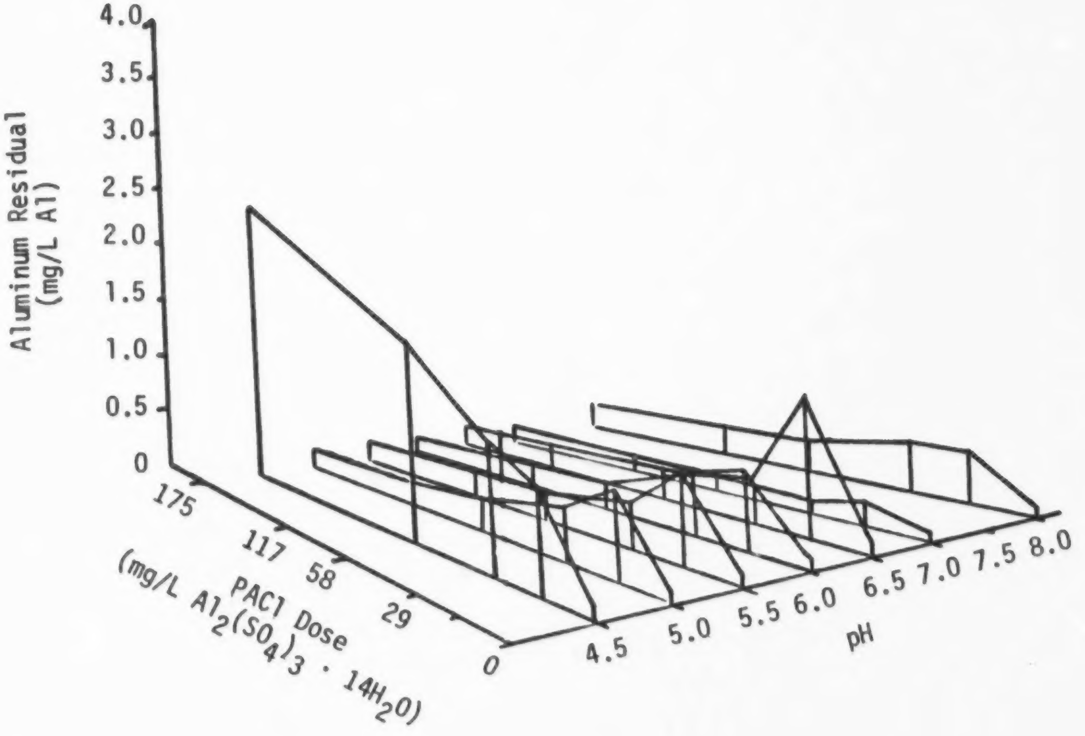


Figure 46 Olga wet season jar tests aluminum residual vs PACl (AC 100S) dose and coagulation pH.

either PACl, and the sulfate-chloride PACl 100S was superior to the chloride PACl 190, for organic reduction. Al residuals were controlled by coagulation pH, being minimized to less than 0.5-mg/L between 5.0 and 7.0. For plant testing the wet season jar tests comparing alum and PACl indicated that pH in the range of 5.0 and 6.0 and an equivalent alum dose of 120 to 240-mg/L $Al_2(SO_4)_3\ H_2O$ was required to investigate organic removal on a plant scale. The intent was to be able to (A) reduce the THM's in the distribution system to 200-300 µg/L, and (B) reduce the color to less than 10 cpu after coagulation and (C) to maintain Al residuals below 0.5-mg/L. These objectives based on bench scale data seemed feasible.

6.2.6 Dry Season Plant Test

The purpose of the dry season plant test was to compare the effectiveness of PACl and alum for the removal of THM precursors during a period of low rainfall. These coagulants were also compared for color, DOC, and TOX precursor removal, Al residuals and sludge volume produced. However, hardness increased as the organics decreased in the raw water during the dry season. Lee County and UCF personnel then made the decision based on raw water quality to soften the water as well as coagulate it in separate reactors. The initial design of the plant incorporated two reactors for the purpose of sequentially softening and coagulation. This effort provided a better water quality for consumers and additional data on organic removal.

The actual dry season testing began on February 7 and continued to March 17. The actual raw water quality during this period is shown in Figure 47. This water is classified as a high color low turbidity water. The turbidity was always less than 4 NTUs and the color ranged from 50 to 60 cpu. It is more difficult to treat than the wet season water because of the high hardness. The hardness averaged approximately 280-mg/L $CaCO_3$ which contained 80-mg/L noncarbonate hardness. The magnesium hardness is not shown in Figure 47 but ranged from 30 to 60-mg/L $CaCO_3$. The higher dry season pH relative to the wet season (8.0 vs. 7.0) is indicative of more groundwater flow into the river during the test. The dry season water would also be treatable by sweep coagulation as color, not turbidity removal, would be the treatment objective. However, because the alkalinity is high, the alum dose required to reduce pH to an acceptable coagulation range is higher than the alum dose necessary for color removal.

Consequently, H_2SO_4 could be added in lieu of alum for pH adjustment at a cost savings. The cost/meq or \$ per H^+ generated is \$0.079 for alum and \$0.034 for H_2SO_4. From the dry season jar test data 85% color removal is possible at a dose of 58-mg/L alum at pH 5.5. However, at typical dry season alkalinity and pH values (160-mg/L and 8) about 85% of that alkalinity must be neutralized to reach pH 5.5 which requires at least 2.7 meq of alum (267-mg/L). If H_2SO_4 were used the cost saving would be \$0.08/1000 gal. However, using alum with acid is not well understood by many operators and this is why so few plants feed acid with alum. The temporary acid feed facilities at Olga are shown in Figure A-5.

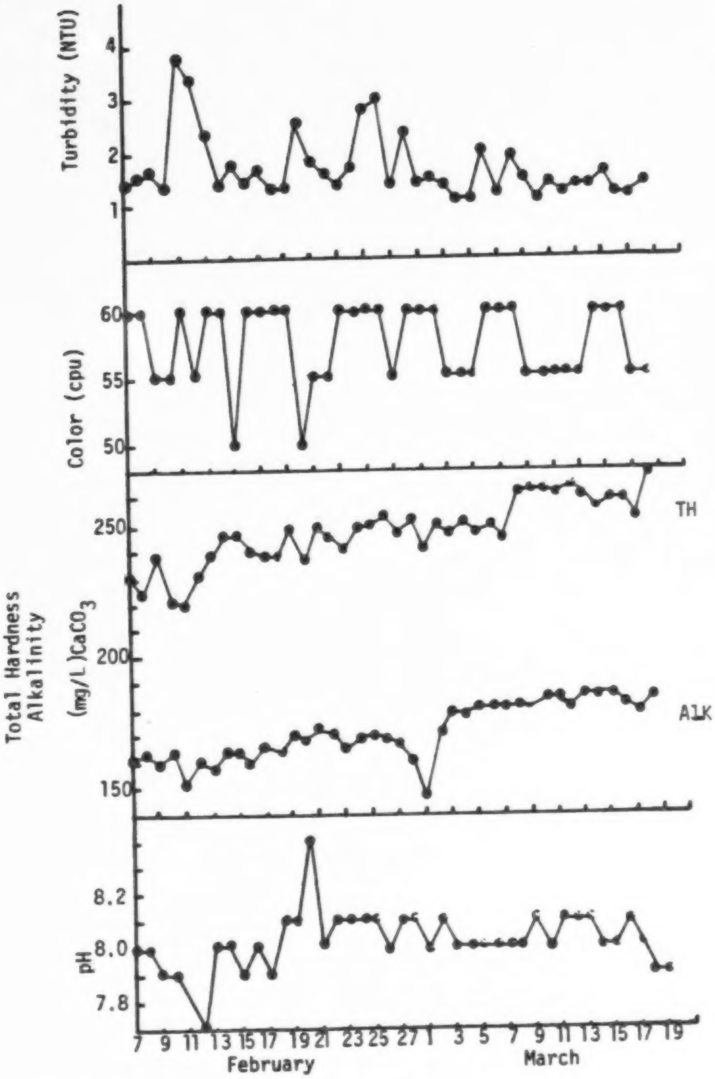


Figure 47 Dry season, raw water, turbidity, color, alkalinity & pH.

The order of softening and coagulation was determined on the basis of cost since earlier data had shown no difference in treatment based on sequence. Softening was done prior to coagulation because the raw alkalinity would be lost if the pH was adjusted to 5.5 to 6.0 prior to softening as required by coagulation. The alkalinity would have to be added to the coagulated water in order to achieve softening which would be expensive. Moreover this sequence would afford an opportunity to demonstrate the lower base neutralizing capacity of PACl since softening would reduce the alkalinity so that the optimum alum dose could not be used without alkalinity addition during coagulation. Since most of the carbonate alkalinity would be lost during softening, a higher PACl dose could be used to reduce the pH to between 5.5 and 6.0 than alum and, therefore, might accomplish more organic removal. The results from the dry season operation are somewhat more complex than the wet season results because chemical additions and water quality are discussed in reference to softening and coagulation as opposed to only coagulation.

The dry season Olga plant demonstration of PACl and alum coagulation started on February 7, 1985. The plant flow diagram for dry season operation is shown in the wet season discussion (see Figure 60) so that seasonal plant operation can be compared. At that time alum feed to the lime-softening unit was discontinued and alum coagulation in the #2 accelerator was commenced. In the initial weeks of the test, pH control in the #2 accelerator was hampered by accumulated lime deposits and mechanical problems with the sulfuric acid pumps. The #2 accelerator was drained and some physically soft sludge was removed during the period of February 16-18. As a result, data were not obtained on those days. The cleaning of the reactor is shown in Figure A-6.

On February 27, sulfuric acid feed problems were partially solved by the installation of a larger pump, and the plant experiments began. The sequence of plant operations for the dry period was:

1. Lime and polymer addition in No. 1 reactor for softening
2. Aeration
3. H_2SO_4 addition for pH adjustment if necessary
4. Alum/PACl addition for organic removal
5. Sodium Hexametaphoshpate addition for stabilization
6. Filtration
7. Chlorination
8. Soda ash addition for final stabilization

The chemical doses and pH used in the softening reactor were initially set to obtain softening to 100-mg/L $CaCO_3$. Typically this means adding CaO until the pH is 10.3 and insuring that the added and raw alkalinity are sufficient to remove the desired calcium hardness. The alum dose in the coagulation reactor was to be used to vary the coagulation pH from 5.5 to 6.5 however additional acid had to be used to lower the pH to a satisfactory value. This was because of hard $CaCO_3$ scale in the reactor that could not be cleaned by the plant personnel. This process sequence is different from the wet season but allowed a comparison of the processes for organic removal. The chemical doses, softening pH, DOC and color data from the softening process for the dry season are presented in Table 45. The chemical doses, coagulation pH, DOC, color and aluminum data from PACl/Alum coagulation process are also presented in Table 45.

6.2.6.1 Softening

As shown in Figure 47, the total hardness (250-mg/L) exceeded the alkalinity (170-mg/L) and in the raw water. A daily average dose of 57-mg/L of alkalinity was added to the raw water to remove the NCH during softening as shown in Figure A-7. The softening pH was varied from 10.6 to 9.5 in the plant test. It was found through in plant jar testing that maximum hardness removal was obtained at pH 9.8 to 10, soda ash additions of more than 60-mg/L removed little if any additional hardness and the CaH could not be reduced to less than 80-mg/L.

The settled TH, CaH, alkalinity, DOC, color and reaction pH of the softening reactor are shown in Figure 48. The addition of alkalinity did reduce the CaH to an average of approximately 100-mg/L. The magnesium hardness (MgH) averaged approximately 40-mg/L. However, in most waters balancing the alkalinity and CaH (both were 210-mg/L after alkalinity addition) should result in a final CaH hardness of 40 to 60-mg/L after softening if the reactor pH is 10.3. This was not the case at Olga and could be due to the organics in the raw water. This would appear to be truly permanent hardness. Additionally the plant data does show on February 10, 11, and 12 softening pHs of 10.4, 10.6, and 10.4 resulted in no additional hardness removal. The addition of approximately 1.2 meq/L of alkalinity could remove an average 0.5 meq/L or 25-mg/L $CaCO_3$ of NCH from this water.

6.2.6.2 Dry Season Plant DOC

During the dry season plant demonstration samples for DOC analysis were taken of the raw water, lime-softening effluent (accelerator #1), coagulated effluent (accelerator #2), filtered water (transfer basin), and the final product at the laboratory tap. The results of these analyses are given in Figure 49.

DOC in the raw water varied from 16.7 to 21.6-mg/L. Lime softening reduced DOC to between 14 and 18-mg/L. If alum was added to the raw water before softening DOC could be reduced to 11-mg/L in the softener effluent. Alum was added on February 7 per normal plant operation for color removal and to aid in settling in the lime-softening unit. During the majority of the test period, however, alum was not used in the lime softener, and DOC reduction was in the range of 11 to 30 percent averaging 20% for accelator #1. After start up the highest DOC removals corresponded to the highest softening pHs, 30% at pH 10.6. The settled DOC reduction after alum coagulation ranged form 13% to 53% after start up based on the incoming or softened DOC. The difficulty in achieving a low pH limited the DOC reduction in reactor #2. This was improved after cleaning but the coagulation pH could be reduced no lower than 6.3 for alum. The average increase in total hardness after coagulation was 20-mg/L $CaCO_3$ due to the dissolution of the hard $CaCO_3$ scale.

TABLE 45. DOC and Color Remaining After Lime Softening During the Dry Season and DOC, Color and Al Residual Remaining after Alum or PACl 100S Coagulation During Dry Season

Date	CaO	Na_2CO_3		DOC		Color	
Date	(mg/L)	(mg/L)	pH	mg/L	%R*	cpu	%R*
Feb. 7, 1985	150	---	9.8	11.0	44	18	72
" 8, "	150	---	9.9	---	---	---	---
" 9, "	150	---	9.9	14.5	21	25	61
" 10, "	149	---	10.4	13.1	33	19	71
" 11, "	143	57	10.6	14.1	29	44	35
" 12, "	140	57	10.4	---	---	---	---
" 13, "	136	57	10.0	14.2	15	39	38
" 14, "	102	57	10.0	---	---	---	---
" 15, "	113	57	10.0	13.4	34	33	48
" 16, "	130	57	9.9	---	---	---	---
" 17, "	150	57	9.9	---	---	---	---
" 18, "	150	57	9.0	---	---	---	---
" 19, "	122	57	8.8	14.5	30	25	58
" 20, "	114	57	10.0	---	---	---	---
" 21, "	96	57	9.5	16.9	17	28	55
" 22, "	110	57	9.8	16.0	29	30	49
" 23, "	110	57	9.9	15.8	24	29	54
" 24, "	110	57	10.0	16.0	20	33	43
" 25, "	110	57	9.9	15.4	27	29	52
" 26, "	110	57	9.9	14.2	28	40	30
" 27, "	110	57	9.8	16.0	20	40	30
" 28, "	110	57	9.8	15.9	19	36	36
March 1, 1985	110	57	9.8	15.8	21	34	39
" 2, "	110	57	9.9	15.3	15	32	46
" 3, "	110	57	9.9	15.7	11	33	43
" 4, "	110	57	9.9	15.8	12	30	56
" 5, "	110	57	10.0	14.9	24	27	51
" 6, "	110	57	9.7	16.7	17	35	36
" 7, "	37	57	9.7	---	---	---	---
" 8, "		---	---	---	---	---	---
" 9, "	88	57	9.7	15.5	14	28	48
" 10, "	100	57	9.7	16.9	16	32	38
" 11, "	100	57	9.7	16.5	16	33	41
" 12, "	100	57	9.8	17.2	18	38	32
" 13, "	100	57	9.9	18.2	16	35	35
" 14, "	100	57	9.7	16.3	21	38	28
" 15, "	101	57	9.5	15.9	24	28	48
" 16, "	110	57	9.5	14.4	25	25	53
" 17, "	110	57	9.6	15.3	15	21	58

*% Removal based on Raw Value

TABLE 45 (Cont.)

	Coag	Dose		DOC			Color			Al
Date	Type	(mg/L)	pH	mg/L	%R_1*	%R_2**	cpu	%R_1*	%R_2**	(mg/L)
Feb. 7, 1985	---	---	9.8	11.5	-5	41	11	39	83	---
" 8, "	---	---	---	---	---	---	---	---	---	---
" 9, "	Alum	185	7.2	12.3	15	33	19	26	71	1.8
" 10, "	Alum	185	8.2	10.3	21	47	11	45	84	1.7
" 11, "	Alum	185	8.2	11.7	17	41	15	67	79	0.6
" 12, "	Alum	---	---	---	---	---	---	---	---	---
" 13, "	Alum	185	7.7	10.1	29	40	11	72	83	0.5
" 14, "	Alum	---	---	---	---	---	---	---	---	---
" 15, "	Alum	185	6.8	10.7	20	47	12	64	81	0.1
" 16, "	Alum	---	---	---	---	---	---	---	---	---
" 17, "	Alum	---	---	---	---	---	---	---	---	---
" 18, "	Alum	---	---	---	---	---	---	---	---	---
" 19, "	Alum	120	6.8	8.1	44	61	5	80	92	1.7
" 20, "	Alum	---	---	---	---	---	---	---	---	---
" 21, "	Alum	250	6.6	8.0	53	61	6	80	91	0.1
" 22, "	Alum	165	6.4	9.5	41	53	7	77	88	0.1
" 23, "	Alum	165	6.7	9.4	41	55	8	72	87	0.1
" 24, "	Alum	165	6.8	9.5	41	53	9	73	84	0.1
" 25, "	Alum	165	6.8	11.4	26	46	12	59	80	0.1
" 26, "	Alum	165	6.7	8.7	39	56	8	82	87	0.1
" 27, "	Alum	120	6.3	9.8	39	51	11	74	82	0.1
" 28, "	Alum	120	6.4	9.7	39	51	8	78	86	0.1
March 1, 1985	Alum	120	6.3	10.0	37	50	8	76	86	<0.1
" 2, "	Alum	80	6.4	11.1	28	38	15	55	75	<0.1
" 3, "	Alum	80	6.7	13.7	13	23	16	52	72	<0.1
" 4, "	Alum	80	6.7	11.3	28	37	11	63	80	0.1
" 5, "	Alum	100	6.7	11.2	25	43	13	52	76	0.1
" 6, "	Alum	100	6.5	11.1	34	45	12	66	78	<0.1
" 7, "	Alum	---	---	---	---	---	---	---	---	---
" 8, "	Alum	---	---	---	---	---	---	--	---	---
" 9, "	PACl	100	6.4	10.6	32	41	10	64	81	0.1
" 10, "	PACl	100	6.8	11.9	30	41	17	47	67	<0.1
" 11, "	PACl	100	6.8	10.5	36	46	11	67	80	<0.1
" 12, "	PACl	120	6.8	15.0	13	28	15	61	73	0.1
" 13, "	PACl	120	7.0	12.9	29	40	15	57	72	0.3
" 14, "	PACl	120	7.5	13.2	19	36	17	55	68	0.2
" 15, "	PACl	80	6.9	11.9	25	43	14	50	54	0.1
" 16, "	PACl	80	6.6	14.4	0	25	8	68	85	0.1
" 17, "	PACl	80	6.0	8.3	46	54	6	72	88	<0.1

* %R_1 is removal based on influent from softening (Acc. #1)

** %R_2 is removal based on raw water value.

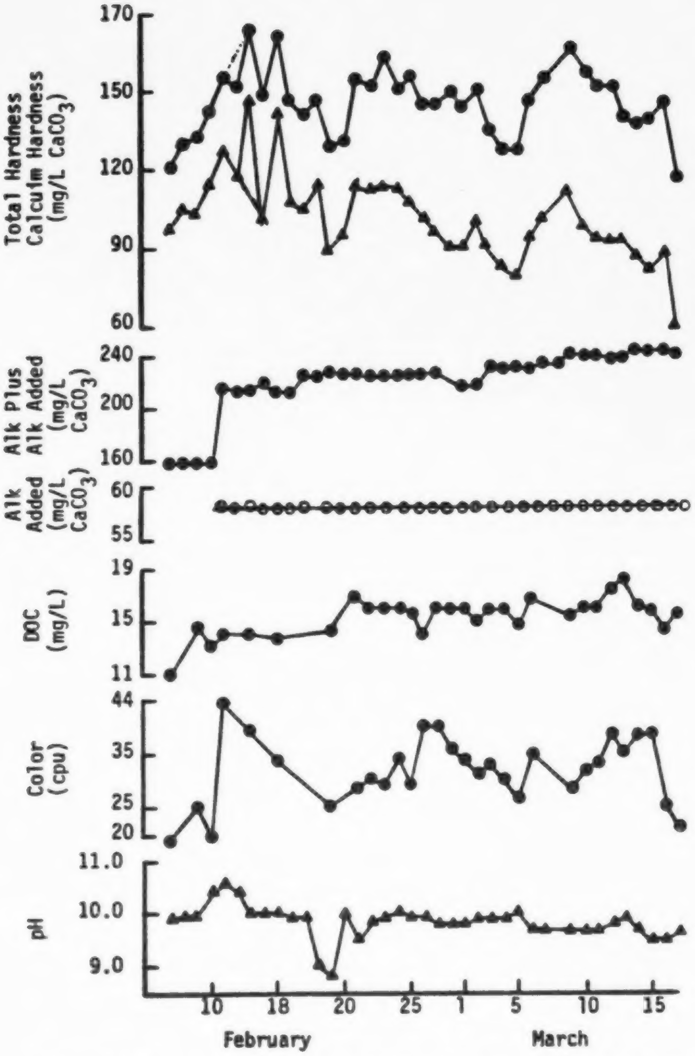


Figure 48 Settled Water Quality from the softening reactor during the dry season.

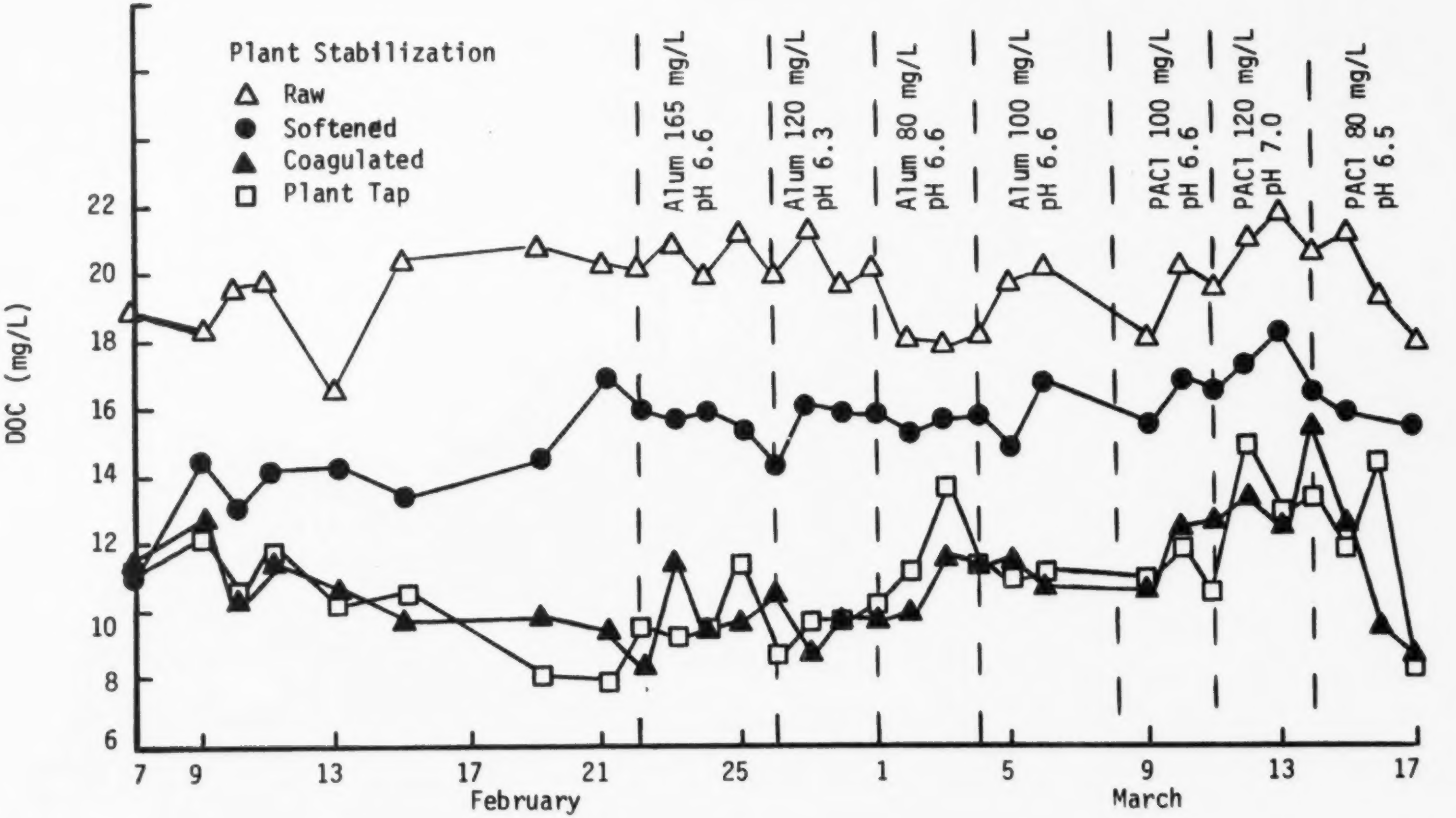


Figure 49 DOC values for raw, softened, coagulated, and plant tap during dry season plant study.

The average DOC percent removal, concentration remaining, coagulant dose and pH are presented in Table 46. These data show that the coagulation pH varied little (6.3 to 6.7) for alum. Varying alum dose to 80, 100 and 120-mg/L at these pHs resulted in a general increase (8 to 13%) for DOC removal at each dose. No significant DOC removal occurs at a 165-mg/L alum dose which indicates removal at that pH was maximized. The removal based on the raw water DOC at alum doses of 80, 100, and 120-mg/L average 50%, 44%, and 33% respectively. The dry season jar tests for removals of DOC should range from 40% to 55% from alum coagulation. The DOC data indicates that preceding alum coagulation with lime softening removed no additional DOC.

The same DOC data analysis for PACl is shown for the dry season in Table 46. It was difficult to lower the coagulation pH with PACl because of PACl's lower base neutralizing capacity relative to alum. Consequently the 120-mg/L dose only removed 22% of the softened DOC for an overall removal of 34%. The 80 and 100-mg/L PACl doses corresponded to 20% and 33% softened DOC removal and 41% and 43% overall. The jar test data for the dry season indicates 30% to 50% overall removal at these pH and dose values. The dry season PACl plant data indicates that preceding alum coagulation with lime softening removed no additional DOC.

The average DOC concentration resulting from treatment with various coagulant doses are given in Table 46. Plotting percent DOC reduction vs. coagulation pH (Figure 50) shows that DOC reduction decreases rapidly as coagulant pH increases above 6.0. A similar trend was also noted for the jar tests performed on the wet and dry season samples. Overall DOC reduction varied between 35% and 66%. A 35 percent reduction was achieved with an alum dose of 185-mg/L and pH = 7.2. Under similar conditions in the jar test phase a DOC reduction of between 40 and 50 percent would be expected. Similarly, 66 percent DOC reduction was achieved with a PACl dose of 80-mg/L as $Al_2(SO_4)_3 \cdot 14\ H_2O$ and a pH of 5.7. Due to the differences in coagulation pH during plant tests, differences between alum and PACl coagulation are difficult to observe. The residual DOC decreases linearly with the coagulation pH for the dry season plant tests as shown in Figure 50 for either PACl or alum. These data show that pH was the most significant factor in the removal of DOC during the dry season. The jar test data show that decreasing pH would have removed more DOC, but that lime softening offers no additional DOC removal preceding coagulation.

The complete dry season DOC variation throughout the plant as shown in Figure 49 indicates the DOC was decreased through softening, coagulation and filtration (only slightly). The variation in DOC as shown in Figure 49 is directly related to the raw water DOC. The DOC remaining after alum coagulation was generally lower than that after PACl coagulation due to the lower pH. The inability of either alum or PACl to realize their complete base neutralizing capacity as indicated by jar tests was a serious loss to the planned dry season plant testing. The $CaCO_3$ scale that was present in the piping and the aerators and #2 reactor kept a potential advantage of PACl from being demonstrated. A normal lime softened water at pH 10.3 will have a total alkalinity of 30 to 50-mg/L which would require only about 60 to 80-mg/L of alum to neutralize it to pH 5.5. If a water needed to be

TABLE 46. Average DOC Remaining for Settled Alum/PACl Coagulation Following Lime Softening During the Dry Season

Dose mg/L Alum*	ALUM			PACl		
	% Removal Soften/Raw	DOC mg/L	pH	% Removal Soften/Raw	DOC mg/L	pH
165	40/53	9.8	6.6	---	---	---
120	38/50	9.2	6.3	22/34	13.0	7.1
100	30/44	11.4	6.6	33/42	10.8	6.8
80	23/33	11.0	6.7	20/41	8.7	6.4

*$Al_2(SO_4)_3 \cdot 14H_2O$

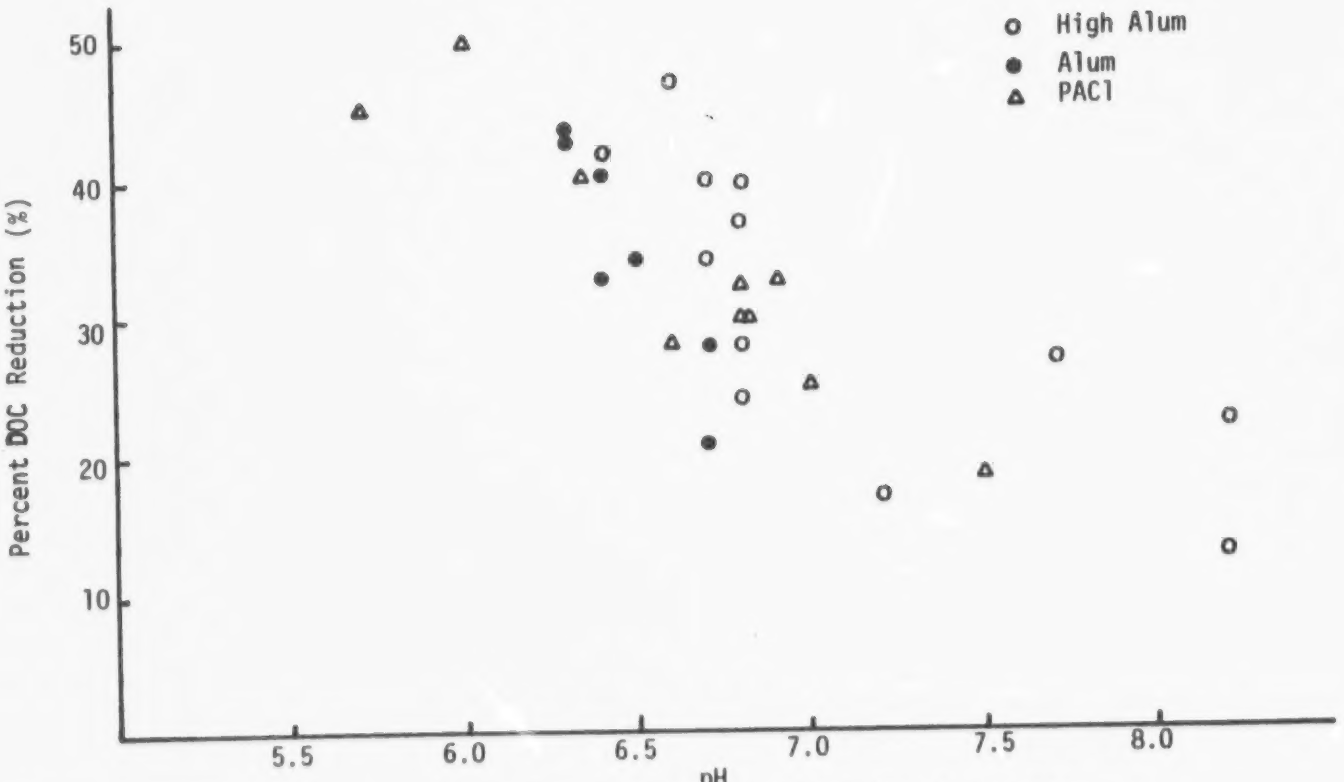


Figure 50 Percent DOC reduction vs coagulation pH during the dry season.

treated for hardness and organic removal a PACl dose of almost twice the alum dose could be used. This can be expanded to any low alkalinity/high color water. In such instances PACl would have a potential advantage over alum. This could have been demonstrated at Olga barring the scale problem.

6.2.6.3 Dry Season THM

During the plant demonstration, trihalomethanes were measured at three points in the plant following chlorination. Samples were taken from the transfer basin immediately following filtration and chlorination, at the inlet to the million gallon storage tank, and at the plant tap. Samples were also taken in the distribution system for analysis both at UCF and at the Lee County Environmental laboratory.

The transfer basin samples represent instantaneous THM formation upon the addition of chlorine to the filtered water. These values ranged from 18 to 100 µg/L. Samples taken at the million gallon tank inlet represented water that had been in contact with free chlorine for a slightly longer time. Since the transfer pumps that supply the storage tank are constantly cycling, it is impossible to say precisely how long the water had been exposed to free chlorine. However it is possible to estimate that the longest period would be no more than 15-20 minutes. THM values ranged from equal to the corresponding transfer basin sample to almost two times the transfer basin concentration. The high degree of variability in values for the sample reflect the cyclical nature of transfer pump operation.

The laboratory sample is actually the first sampling point in the distribution system. Water samples at this point had been in contact with chlorine from 12 to 18 hours. THM concentrations at this point were often only slightly higher than the inlet samples, while at other times the concentrations were 2 to 4 times higher than the tank inlet and from 2 to 5 times higher than the transfer basin. Average values for THM concentration ranged from 174 µg/L for alum at 120-mg/L to 215 µg/L for both alum at 80-mg/L and PACl at 120-mg/L. The Cl_2 dose and THM's at the plant tap are presented in Figure 51. This data shows that THM's at the tap ranged from 65 µg/L to 240 µg/L after plant stabilization. The tap THM's closely follow Cl_2 dose and are about 10% higher when PACl was used which is about the difference between the 8 to 9-mg/L Cl_2 dose corresponding to alum and the 9 to 10-mg/L Cl_2 dose corresponding to PACl treatment. Plant personnel had judged it was more difficult to maintain a Cl_2 residual with PACl than Alum. Although PACl may be slightly less effective for organic removal than alum, the difference in Cl_2 dose would mask any precursor THM removal difference attributable to the coagulants themselves.

During the dry season plant demonstration, distribution system samples were also taken and analyzed by Lee County personnel. A map of those sites is shown in Figure 52. Analyses included trihalomethane and aluminum residuals. Sampling points included seven points in the distribution system and the plant tap. The Lee County distribution system runs west along S.R. 80 toward Fort Meyers for approximately 4 miles to County Road 31. At this point the distribution system splits with most of the water crossing the Caloosahatchee River and a smaller branch continuing west

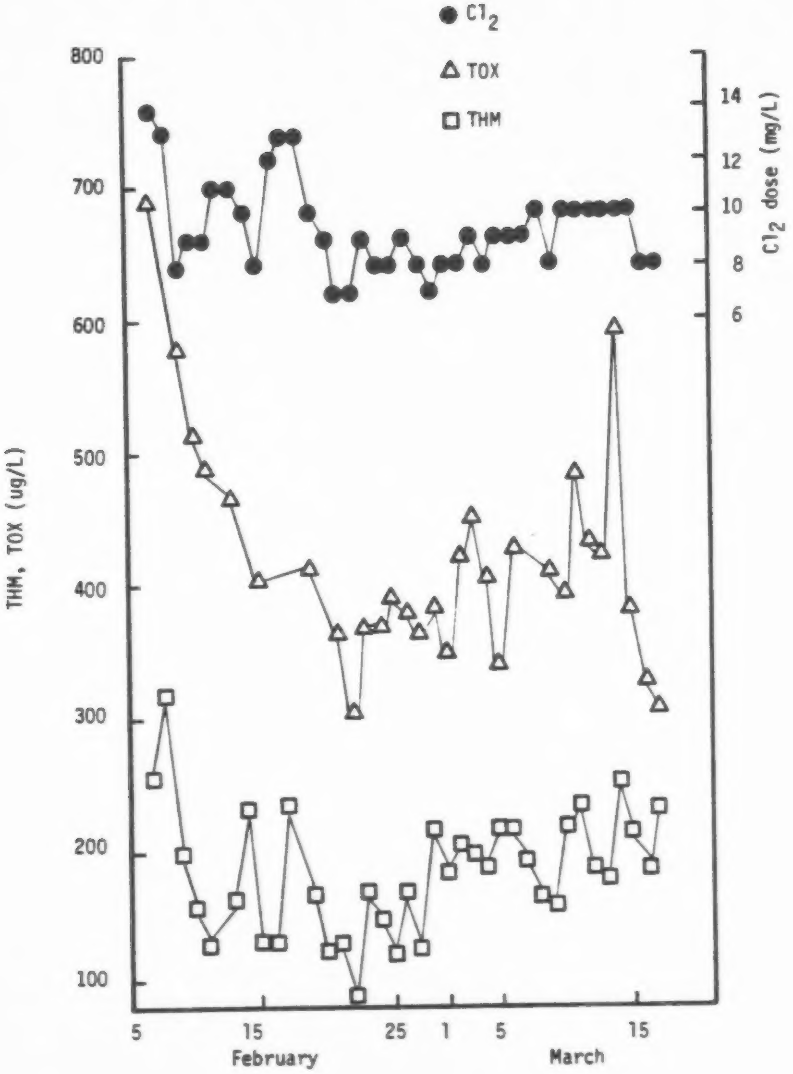
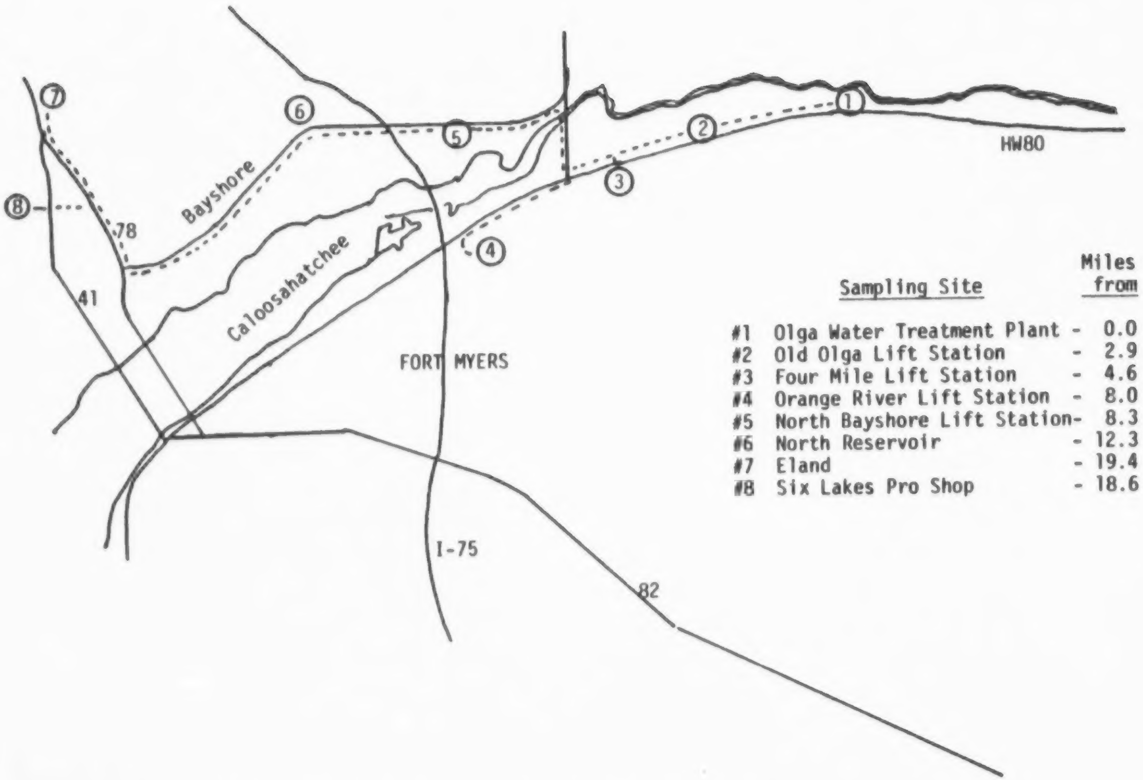


Figure 51 Plant tap THM, Plant tap TOX, and Cl_2 dose during dry season plant study


HW80
Bayshore
Caloosahatchee
FORT MYERS
I-75
82
41
78
①
②
③
④
⑤
⑥
⑦
⑧

Sampling Site	Miles from
#1 Olga Water Treatment Plant	- 0.0
#2 Old Olga Lift Station	- 2.9
#3 Four Mile Lift Station	- 4.6
#4 Orange River Lift Station	- 8.0
#5 North Bayshore Lift Station	- 8.3
#6 North Reservoir	- 12.3
#7 Eland	- 19.4
#8 Six Lakes Pro Shop	- 18.6

Figure 52 Map of Sampling Points in the Olga Distribution System.

along S.R. 80. This line continues for approximately 2 miles further and services residences east of I-75. The branch that crosses the Caloosahatchee River travels west along S.R. 78 north of the river to the North Reservoir, approximately 2 miles west of I-75. Here the water is stored, chlorinated as needed, and pumped to the rest of the distribution system. This extends west to US 41 and approximately 1 mile north on US 41.

THM values at the Old Olga Road Lift Station were almost identical to values at the plant tap as shown in Table 47. By the time the water reached the N. Reservoir, THM concentrations had increased by 100-200 µg/L in some cases. Despite additional chlorination and residence time, THM concentrations at the Eland sampling point were usually very similar to the N. Reservoir values.

Water samples taken at the North Reservoir and Eland both represent samples with significantly longer residence times than the lab tap. Additional chlorine was added at the North Reservoir in order to maintain a residual in the distribution system. THM's for the North Reservoir varied from 200 to 400 µg/L during the study period. From 2/27 to 3/20, roughly the period of the plant testing in the dry season, THM's did not exceed 300 µg/L. THM values at Eland were very similar to those at N. Reservoir. The Eland station was the farthest point in the distribution system and thus should represent the highest THM levels expected. Values at this station ranged from 200 µg/L to over 400 µg/L during the study period. Figure 53 shows THM formation throughout the Olga distribution system. THM formation was approximately 66% complete at the plant tap and nearly 100% complete at the N. Reservoir. No clear differences are detectable in distribution system THM levels that would separate PACl and alum treated water. However, THM levels recorded during the test period were a major improvement over water quality prior to the beginning of alum/PACl coagulation. The THM's in the Olga distribution system are listed in Table 47.

Due to the variability in residence times it is not likely that water produced on individual days of operation could be identified at the various points in the distribution system. In order to compare operation with alum and PACl averages representative of the entire operation for each coagulant were compared as shown in Table 47. Significant differences are noted for the plant samples and those samples taken at the North Reservoir and beyond. In these cases, alum resulted in significantly lower THM values than PACl.

6.2.6.4 Dry Season TOX

Samples for TOX analysis were taken from the raw water, transfer basin, storage tank inlet, plant tap, and in the distribution system from the North Reservoir and Eland Station. The average value for TOX in the raw water was 37 µg/L and varied between 18 and 64 µg/L. A plot of this data is shown in Figure 54. The presence of TOX in the raw water is not surprising since the Caloosahatchee drains a watershed that is predominantly agricultural (sugar cane fields).

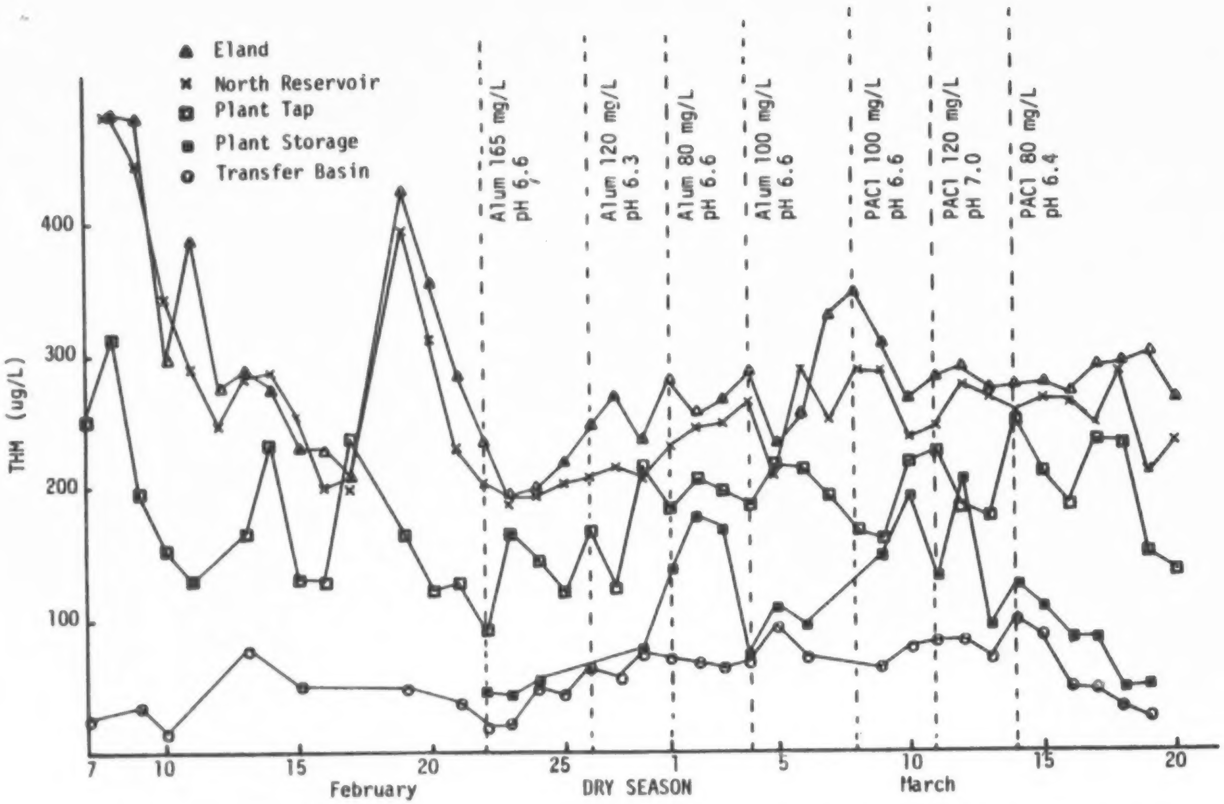


Figure 53 THM Concentrations in the Olga finished water during the dry season test throughout the distribution system and plant.

TABLE 47. Olga Dry Season Plant Tests THM (µg/L) - Distribution System.

Date	1	2	3	4	5	6	7	8
2-8-85	463	454	320	370	482	482	502	316
2-9-85	442	380	246	391	444	479	510	212
2-10-85	476	287	214	275	342	293	415	197
2-11-85	318	261	164	250	296	339	368	154
2-12-85	309	214	172	227	246	274	300	
2-13-85	218	258	191	247	243	249	277	192
2-14-85	292	285	233	269	288	275	295	233
2-15-85	366	242	114	192	254	230	307	96
2-16-85	226	164	156	159	199	231	265	133
2-17-85	196	276	283	233	210	209	241	238
2-18-85								
2-19-85	429	290	275	370	357	391	383	251
2-20-85	337	276	121	234	321	345	388	121
2-21-85	265	242	164	252	229	282	288	164
2-22-85	229	220	160	194	197	236	233	110
2-23-85	205	183	143	190	193	206	242	125
2-24-85	205	197	155	193	212	217	210	144
2-25-85	201	203	152	226	203	175	241	141
2-26-85	206	222	174	199	187	202	156	168
2-27-85	224	233	142	216	190	218	177	123
2-28-85	179	206	206	197	190	204	196	223
3-1-85	244	238	150	231	233	236	239	140
3-3-85	278	275	192	277	202	268	271	161
3-3-85	229	221	195	239	228	243	267	171
3-4-85	240	239	173	263	217	232	252	167
3-5-85	260	244	195	224	207	227	244	157
3-6-85	235	225	210	244	239	240	241	182
3-7-85	264	224	189	233	211	247	273	191
3-8-85	239	259	180	230	240	280	271	163
3-9-85	284	258	198	246	235	266	276	198
3-10-85	254	227	206	237	242	241	276	184
3-11-85	249	283	256	282	232	258	261	217
3-12-85	282	288	238	281	259	258	287	194
3-13-85	319	288	223	282	265	294	276	188
3-14-85	278	277	267	267	255	260	266	219
3-15-85	280	308	218	272	262	309	308	214
3-16-85	291	283	221	255	261	292	285	185
3-17-85	150	289	154	232	257	257	294	232
3-18-85	150	289	154	232	257	257	294	232
3-19-85	241	195	168	168	201	217	1987	177
3-20-85	209	163	148	191	231	255	261	138
3-21-85	253	247	243	259	252	238	246	203
3-22-85	295	360	357	340	324	290	276	296
3-23-85								
3-24-85	37	356	245	301	350	357	392	233

TABLE 47 (Cont.)

Site	1	2	3	4	5	6	7	8
Coagulant								
Alum	171	188	241	253	242	225	248	261
	6	5	6	6	5	5	6	5
PACl	198	199	264	244	242	250	267	284
	10	14	17	21	14	6	9	6

x = Average THM ug/L

s = Standard Deviation ug/L

Site	Location
1	Orange River Lift Station
2	4 M Lift Station
3	Old Olga Lift Station
4	West Bayshore Lift Station
5	N. Reservoir Storage Tank
6	Eland
7	Six Lakes Mobile Home Park
8	Plant Tap

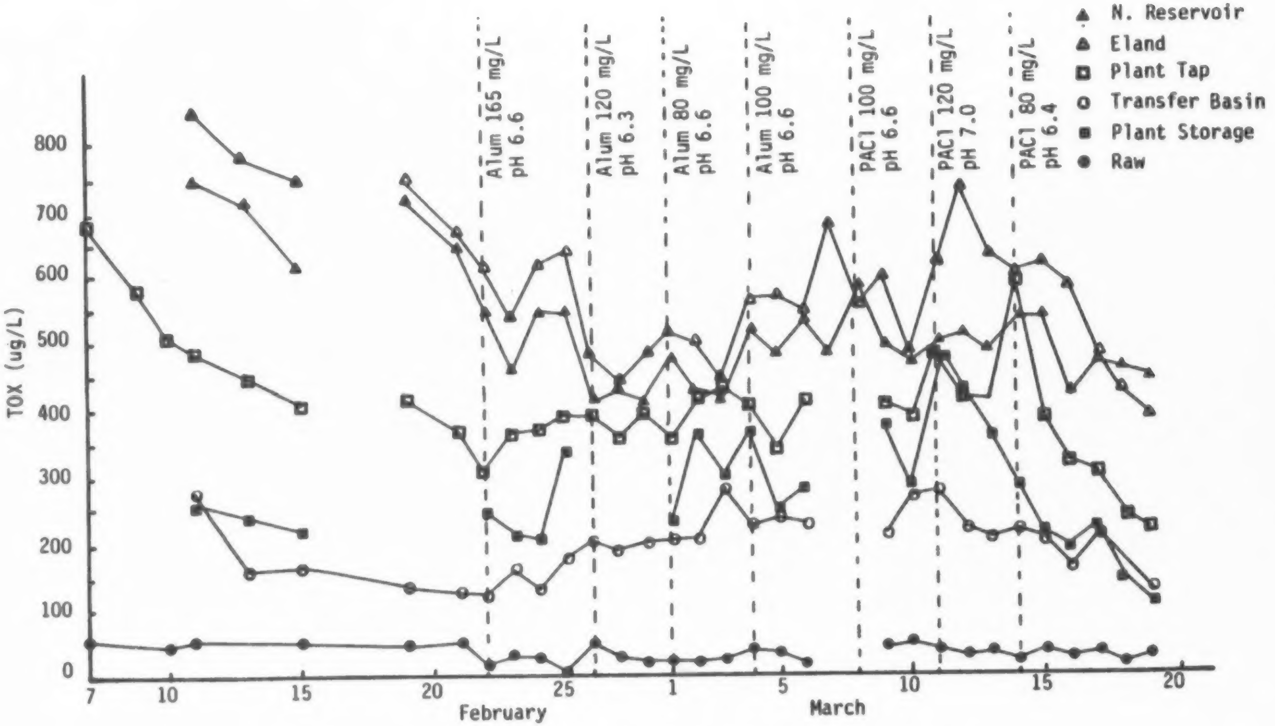


Figure 54 TOX concentrations in the Olga finished water during the dry season tests throughout the distribution system and plant.

TOX values in the transfer basin represent the instantaneous formation of halogented organic compounds upon the addition of chlorine. These values varied between 117 and 284 µg/L and represented approximately 30 to 60% of the TOX concentration measured at the lab tap. As was noted for THM concentrations, TOX varied greatly at the storage tank inlet, probably due to the cycling of the transfer pumps. TOX at the lab tap varied between 231 and 689 µg/L. This high value occurred on the first day of the plant demonstration and is not representative of the coagulated water. After the accelerator #2 was cleaned, TOX values were consistently less than 450 µg/L and averaged 382 µg/L from Feb. 20 to March 17.

As was the case with alum, PACl samples responded more to coagulation pH than to dose for TOXFP removal. As shown in Figure 55, TOX varies inversely with coagulation pH. The data given in Table 48 shows that TOX did decrease with increasing alum dose but increased with PACl dose because of a increasing PACl coagulation pH. This indicates that coagulant doses were adequate to remove the maximum TOX precursors at the pHs utilized. This was the same observation during the dry weather plant tests for DOC and THM precursors. The exact TOX values for Figure 54 are listed in Table 49 with the TOX:THM ratios. The average TOX found in the tap samples after plant testing began (2/22/85) was 382 µg/L. The average TOX found in the Eland distribution system (furthest point) was 546 µg/L which means that 70% of the TOX was formed before the water left plant storage. TOX values varied from 748 to 395 µg/L at the Eland station. Approximately one-half of the TOX formed in the plant (35% of the total) was formed quickly in transfer basin immediately after chlorination. This trend can be seen in Figure 54. A general statement can be made that 1/3 of the total TOX was formed instantaneously in the transfer basin and 1/3 was formed during the plant storage (1 day) and 1/3 was formed in the distribution system. The DOC:THM:TOX ratio for the dry season tap samples is presented in Table 49. The average DOC:THM:TOX ratio is 1:16:39. This ratio varied from 1:17:31 to 1:21:42 after plant stabilization. The TOX:THM ratio was 2.4:1. The dependency of THM on Cl_2 dose can be seen in Figure 56. After plant stabilization the Cl_2 dose ranged from 7-mg/L to 10-mg/L and was slightly higher for PACl coagulation. The increasing of Cl_2 dose from 2/21/85 is accompanied by a corresponding increasing THM formation as shown in Figure 56.

6.2.6.5 Dry Season Color

Color was measured during the dry season testing at five locations in the plant which were the raw water, settled water after softening, settled water after coagulation, transfer basin and the color tap. This data is given in Table A-8 and is shown in Figure 57.

Raw color varied from 68 to 50 cpu. Color removal in the softening reactor ranged from 30% to 58% which represented a settled color of 20 to 40 cpu. The average settled softened color was 32 cpu which was 43% removal. The color data in Figure 56 show that the color removal during softening did vary but generally decreased as the raw color decreased.

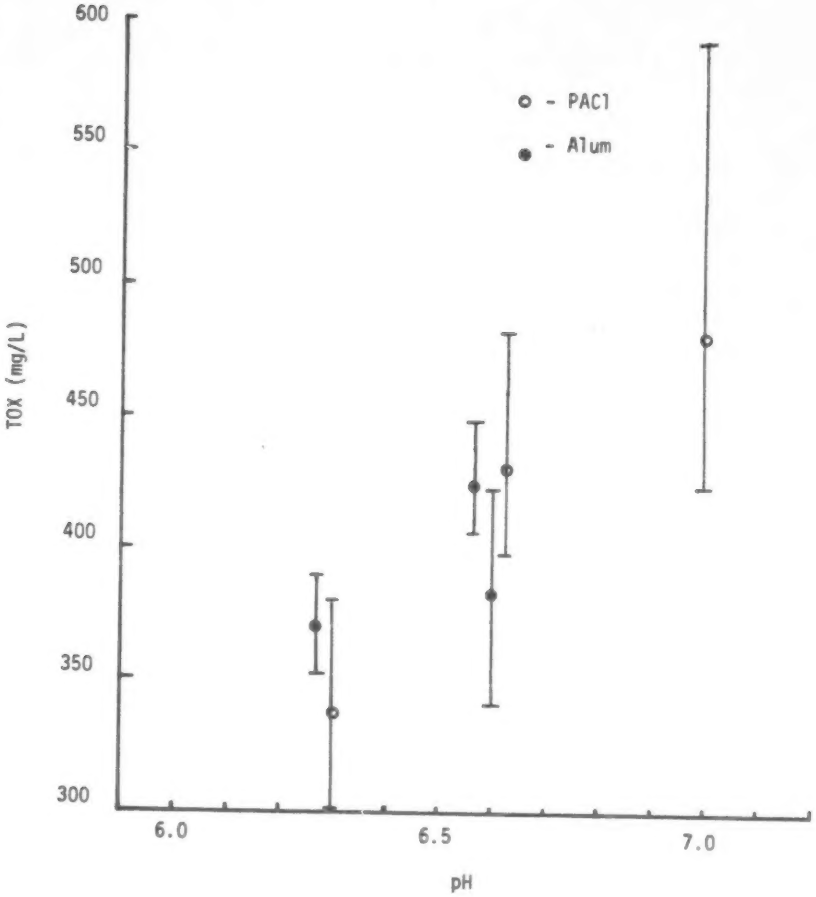


Figure 55 Average finished water TOX vs. average coagulation pH during Olga Water Treatment Plant dry season plant trials.

TABLE 48. Dry Season Average THM by Coagulant Dose as $Al_2(SO_4)_3 \cdot 14H_2O$ and pH

Alum Dose mg/L	pH	THM x ug/L	THM s ug/L
120	6.3	370	11
100	6.6	383	43
80	6.6	425	12

PACl Dose mg/L	pH	THM x ug/L	THM s ug/L
120	7.0	482	54
100	6.6	430	26
80	6.3	338	28

x = mean

s = standard deviation

TABLE 49. DOC, THM and TOX Ratios from the Plant Tap During the Dry Season

DOC (mg/L)	InstTHM (ug/L)	THM (ug/L)	DOC/THM/TOX	TOX/THM
11.6	256	689	1/22/59	2.7
12.8	196	578	1/15/45	3.0
10.3	156	515	1/15/50	3.3
11.4	130	491	1/11/43	3.9
10.5	163	465	1/16/44	2.8
9.7	132	403	1/14/42	3.0
9.9	165	416	1/13/42	3.2
9.3	128	363	1/14/39	2.8
8.4	91	306	1/11/36	3.3
11.4	168	369	1/15/39	2.6
9.4	144	369	1/15/39	2.6
9.7	120	391	1/12/40	3.3
10.4	168	380	1/16/37	2.3
8.7	123	366	1/14/42	3.0
9.7	216	390	1/22/40	1.8
9.6	182	353	1/19/37	1.9
10.0	206	420	1/21/42	2.0
11.5	197	448	1/17/39	2.3
11.1	183	406	1/16/37	2.3
11.3	216	340	1/19/30	1.6
10.7	213	426	1/20/40	2.0
10.4	159	412	1/15/40	2.7
12.3	217	396	1/18/32	1.8
12.6	227	482	1/18/38	2.1
13.2	182	431	1/14/33	2.4
12.5	177	424	1/14/34	2.4
15.3	251	591	1/16/39	2.4
12.4	210	380	1/17/31	1.8
9.4	183	326	1/20/35	1.8
8.5	232	308	1/27/36	1.3
7.0	232	246	1/33/35	1.1
6.2	149	231	1/24/37	1.5
Average	173	410	1/16/39	2.4

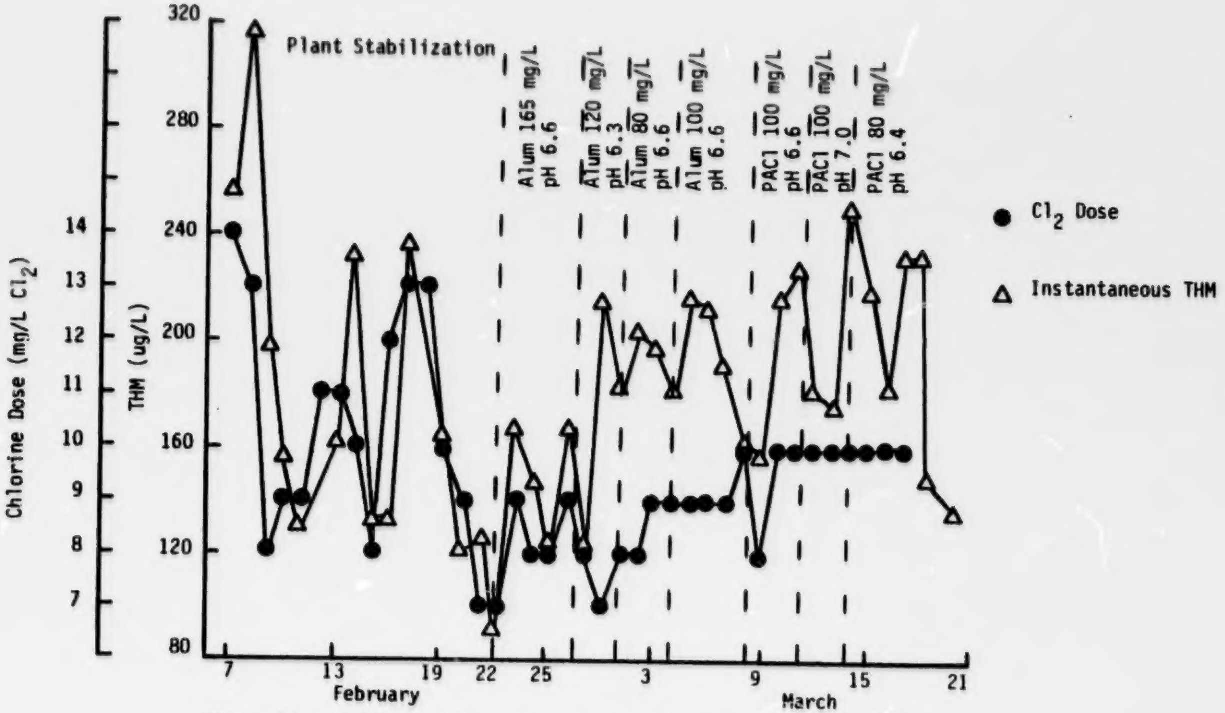


Figure 56 THM and chlorine dose during the dry season plant test.

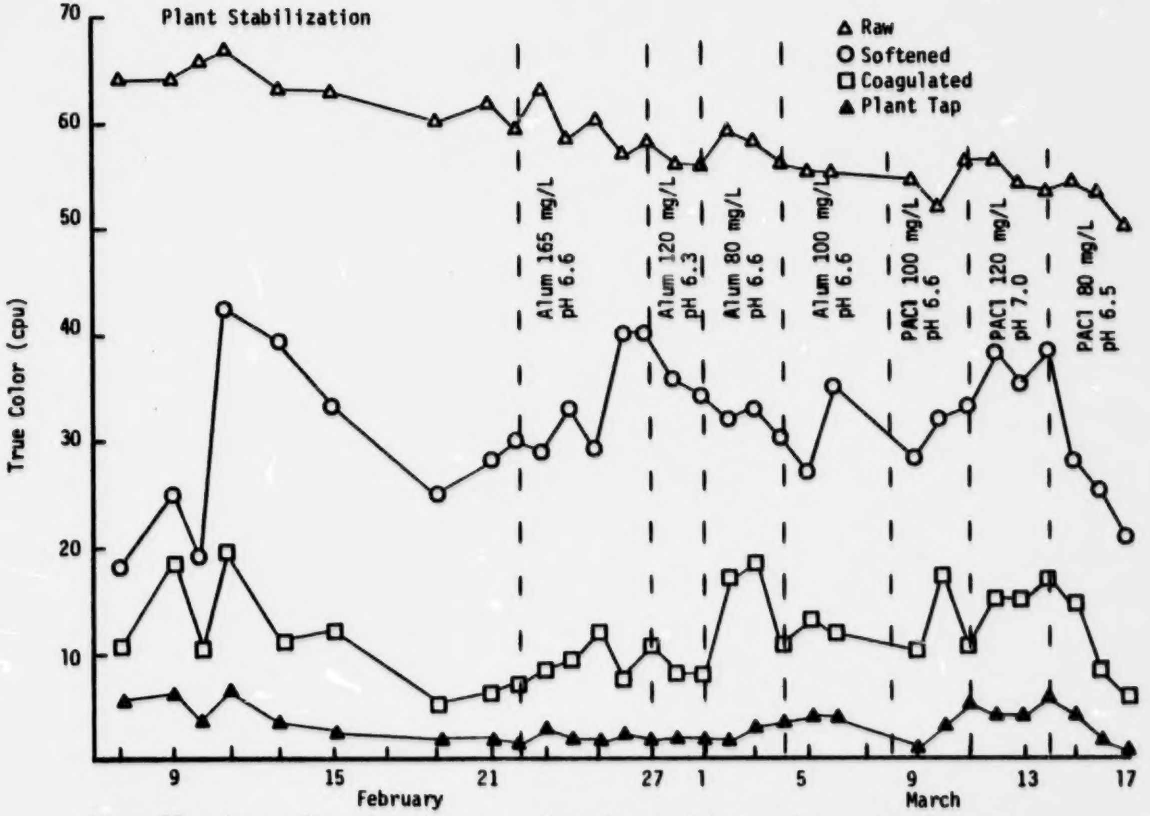


Figure 57 True color values for raw, softened, coagulated, and plant tap during dry season

The color removal during alum or PACl coagulation varied from 52% to 80% for alum and 54% to 81% for PACl. As with DOC, THM, and TOX precursor removal, the removal of color was controlled by coagulation pH. Alum doses of 165, 120, 100, and 80-mg/L averaged 75%, 76%, 59% and 56% color removal. The maximum color removal was with alum at 76% for a dose of 120-mg/L. PACl doses of 120, 100, and 80-mg/L removed 58%, 59% and 63% of the softened color. Variation of PACl coagulation pH reversed the expected PACl color removal results as PACl dose was increased and hindered overall organic removal during coagulation. The overall color removal ranged from 88% to 76% for alum and from 88% to 54% for PACl. As with DOC the earlier dry season jar test show that from coagulation pHs of 6.5 to 7.0 for alum or PACl doses of 80 to 120-mg/L should get 50% to 80+% color removal. This data indicates there is very little advantage to preceding alum or PACl coagulation with softening for color removal.

The variation in raw water, softened, coagulated and tap colors is shown in Figure 57. Contrary to the similar dry season DOC data shown in Figure 48 this figure shows the majority of the total color removal was accomplished during softening, but generally alum or PACl coagulation lowered the settled color to less than the 15 cpu regulation. The color removed by filtration and chlorination ranged from 0.5 to 1.0 cpu/mg/L Cl_2. Filtration removed some color but it was impossible to differentiate from chlorination since chlorination occurred in the transfer basin immediately after filtration.

6.2.6.6 Other Dry Season Parameters

Aluminum residuals were measured by Lee County personnel at the plant tap and at 7 sites in the distribution system. These values are given in Table A-9. Figure 58 shows the Al residual at the plant tap during the dry season. Residuals in the plant effluent were consistently less than 0.20-mg/L during the initial start up, due to the high residuals remaining in the storage tank from the normal operation of the plant. Also during and immediately following the period when the #2 accelerator was being cleaned, aluminum residuals increased above 0.2-mg/L due to the return of the plant to normal operations.

These brief periods of high aluminum residual allow a rough estimation of the residence time within the Olga distribution system. From the data, it can be seen that the samples from the Old Olga Road Lift Station followed very closely the fluctuations of aluminum residual in the plant effluent. The Orange River Lift Station, the farthest part of the distribution system on the south side of the Caloosahatchee appear to have a 1 to 1.5 day delay in responding to changes in water quality in the plant effluent. On the north side of the river, the North Reservoir had a 1 to 1.5 day delay relative to the plant and the sampling point at Eland lagged approximately 2 to 3 days behind the plant effluent. Water quality fluctuation were somewhat dampened by the storage capacity of the North Reservoir. The total residence time in the distribution system including plant storage probably did not exceed five days.

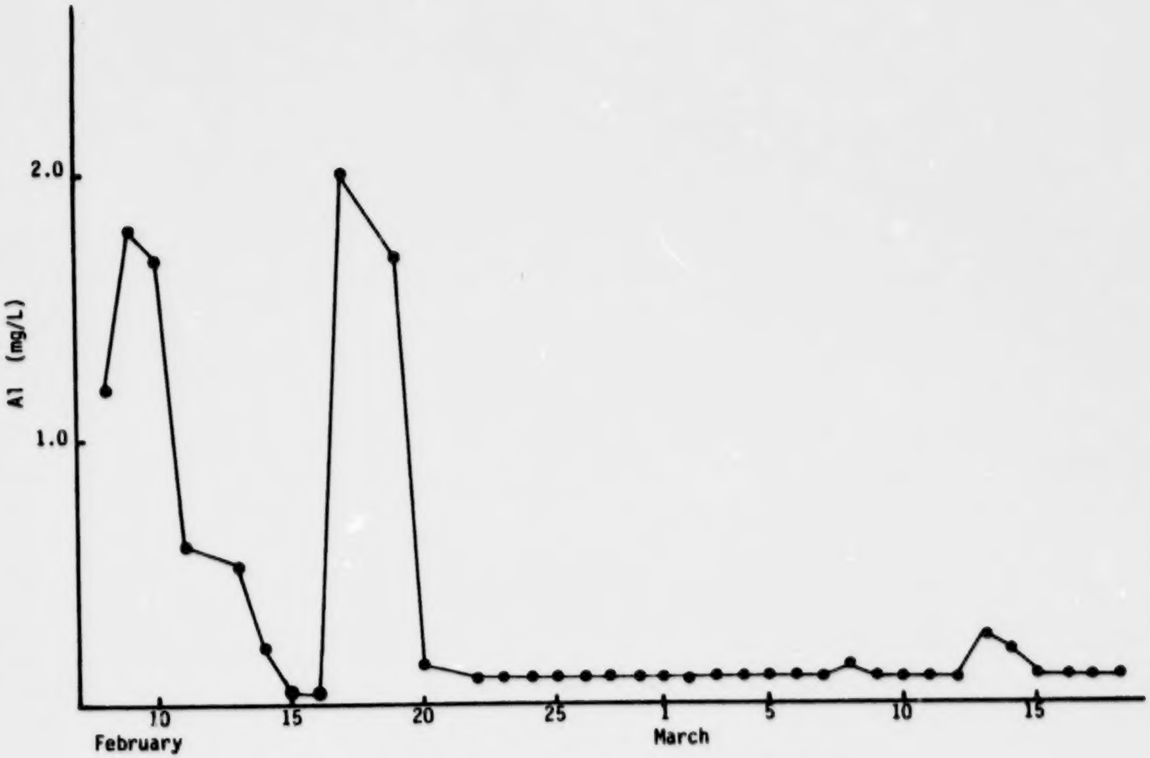


Figure 58 Al residuals at the plant tap during the dry season.

Aluminum residuals for operation with PACl and alum were essentially identical. The only significant increase in aluminum concentration occurred on March 13 and 14 when pH in the coagulation basin increased to over 7. On these days aluminum residuals increased to over 0.2-mg/L. These values are consistent with aluminum residuals found in jar tests conducted at these pH values. Residuals during the rainy season plant demonstration averaged 0.2-mg/L. No significant difference was noted between PACl and alum Al residuals in either case.

During the dry season the plant was operated at 2000 gpm almost 24 hrs/day as the demand was higher. The surface loading rates on the reactors was 0.53 gpm/ft^2 which is well below the 1.0 gpm/ft^2 design rate for $CaCO_3$ settling and approximately the same as the 0.50 gpm/ft^2 design rate. As in the wet season no hydraulic stress was placed on either the softening or the coagulation reactor or the filter during plant tests.

The dry season turbidities for the softening reactor, coagulation reactor and finished water are shown in Table 50. There was a higher carry over turbidity, approximately 10 NTU, from the softener than desired. The additional soda ash feed did form some very fine $CaCO_3$ floc which the plant polymer alone could not settle. Alum was usually fed in the raw water lines but could not be at this time because the small membrane pilot plant was being fed from this source. As in the dry season the turbidities resulting from either PACl or alum coagulation were essentially the same. The turbidities, however, were 1 to 2 NTU's on average higher than the corresponding wet season. This was due to the hard $CaCO_3$ scale that could not be quickly removed, but was slowly dissolving at coagulation pHs of 5.0 to 7.0 and causing some turbidity.

The sludge volume produced after 10 minutes of settling of a one liter sample collected from the center (rapid mix.) of the coagulation reactors is given in Table 51. These data show the alum sludge samples averaged 49% after settling and had a dose/percent volume ratio of 2. Although this test is designed to control sludge volume in a reactor it indicates that there is no difference in the PACl or alum sludge volume produced during the dry season tests.

6.2.6.7 Dry Season Summary

The dry season plant testing included lime softening prior to PACl or alum coagulation. The primary purpose of lime softening was hardness removal although DOC and color removal averaged 20% and 43% removal. Approximately 1.2 meg/L of alkalinity was added to the raw water to minimize the calcium hardness. The final calcium hardness could only be reduced to approximately 100-mg/L which would have been reduced to 50-mg/L or less in conventional Florida ground waters (low organic).

The hard $CaCO_3$ scale in the plant piping and reactor 2 kept the alum/PACl coagulation pH higher than desired, 6.5 to 7.0 vs 6.0 to 5.5. Acid feed was required to reduce the pH to an acceptable range. The artificially higher pH kept the lower base neutralizing capacity of PACl from being demonstrated to potential advantage for organic removal.

TABLE 50. Olga Dry Season Turbidity

Date	Softening NTU	Coagulation NTU	Finished NTU
2/7*	4.0	2.9	0.2
2/8	5.0	3.3	0.2
2/9	13.4	2.2	0.3
2/10	20.3	2.1	1.5
2/11	8.7	0.9	1.1
2/12	10.2	0.8	0.9
2/13	17.0	1.4	0.3
2/14	11.3	0.7	0.1
2/15	19.3	2.7	0.2
2/16	27.5	3.2	0.1
2/17	4.7	---	0.7
2/18	4.7	5.7	0.3
2/19	1.8	6.9	0.4
2/20	6.5	0.9	0.3
2/21	14.0	1.2	0.1
2/22	5.7	1.2	0.1
2/23	10.0	1.9	0.1
2/24	6.8	1.5	0.2
2/25	6.2	1.9	0.1
2/26	8.2	1.9	0.2
2/27	8.3	1.6	0.2
2/28	11.7	2.0	0.2
3/1	13.9	1.7	0.2
3/2	13.4	1.8	0.2
3/3	11.4	2.0	0.2
3/4	12.4	2.1	0.3
3/5	11.9	2.5	0.4
3/6	11.1	1.8	0.1
3/7	38.0	1.9	0.2
3/8	---	1.4	0.1
3/9**	4.1	1.6	0.2
3/10	7.2	1.6	0.2
3/11	8.5	1.9	0.3
3/12	7.1	1.9	0.3
3/13	13.3	1.9	0.3
3/14	16.4	1.8	0.3
3/15	14.4	2.0	0.2
3/16	10.0	1.4	0.3

*: Alum Coagulation to 3/8
**: PACl Coagulation to 3/16

TABLE 51. Settled Sludge Volume for Dry Season Plant Test

Date	Coagulant Dose (mg/L)	Volume* (%)	Dose/Volume%
Feb. 11, 1985	Alum 185	34	3
" 12, "	" 120	52	2
" 13, "	" 185	53	3
" 14, "	" 185	54	5
" 15, "	" 185	37	4
" 16, "	" 185	42	3
" 17, "	" 65	25	2
" 18, "	" 65	28	4
" 19, "	" 120	32	7
" 20, "	" 250	35	5
" 21, "	" 250	48	4
" 22, "	" 165	42	4
" 23, "	" 165	40	3
" 24, "	" 165	52	4
" 25, "	" 165	44	4
" 26, "	" 165	42	3
" 27, "	" 120	47	3
" 28, "	" 120	45	4
March 1, 1985	" 120	48	3
" 2, "	" 80	45	2
" 3, "	" 80	42	2
" 4, "	" 80	46	2
" 5, "	" 100	42	2
" 6, "	" 100	47	2
" 7, "	" 100	47	2
" 8, "	" 100	51	2
" 9, "	PACl 100	49	2
" 10, "	" 100	50	2
" 11, "	" 100	48	2
" 12, "	" 120	49	2
" 13, "	" 120	51	2
" 14, "	" 120	49	2
" 15, "	" 80	53	2
" 16, "	" 80	48	2
" 17, "	" 80	42	2
" 18, "	" 80	48	2

*Percent volume settled sludge occupies after 5 minutes of settling in cone.

The dry season jar tests had indicated that color and DOC removals at alum or PACl doses of 80, 100, and 120-mg/L as alum should remove between 30% to 55% of the DOC and 60 to 80% of the color. The DOC and color removal based on the raw water was in this range for DOC and slightly better (5-10%) for color during the plant test. The results indicate that preceding alum or PACl coagulation with lime softening at pH's of 9.5 to 10.5 does not increase organic removal even though the organics (after softening) were reduced (20% DOC and 43% color on average). The jar test data indicates the DOC removal from the plant could have been increased by approximately 5 to 10% if the plant coagulation pH could have been lowered to 5.5 which would have also lowered plant THM and TOX. DOC removal by alum was higher than by PACl however this was more effected by coagulation pH than by dose. Increasing alum doses resulted in increasing DOC removal whereas the opposite was true for PACl because of a higher coagulation pH. But both the alum and PACl DOC removals were within the jar test predictions which indicated alum removes slightly more DOC than PACl at equivalent conditions.

THM's in the system ranged from 65 to 240 μg/L at the plant tap to 200 to 400 μg/L at the extremities of the distribution system. As with DOC the lowest plant tap THM's were found after alum coagulation, but this was probably more attributable to the higher Cl_2 dose used during PACl coagulation than any other factor.

TOX formation varied from approximately 600 to 300 μg/L during the dry season plant test. The TOX formed varied directly with the THM formed which were both directly dependent on Cl_2 dose. The average ratio of DOC:THM:TOX was 1:16:39 and the average THM:TOX ratio was 2.4:1. For given chlorination conditions (dose, pH, temperature, residence time, DOC) TOX was somewhat predictable by THM.

No difference between the Al residuals, sludge produced or turbidity carry over to the filters was observed during the dry season test for PACl or alum coagulation. Al residuals were always less than 0.5-mg/L Al for the dry season tests.

Although the jar test data shows a slight (5%) increase in color or DOC removal by alum vs PACl coagulation the practical significance of this fact in terms of plant operation is very little. The plant data shows for the dry season treatment of a low turbidity high color water that PACl offers no advantage over alum for organic removal.

6.2.7 Wet Season Plant Test

6.2.7.1 Introduction

During the wet season in July and August, 1984, and during the dry season in February and March, 1985, PACl coagulation was compared with conventional alum coagulation at the Olga Lee County Water Treatment Plant, located near Fort Meyers, Florida. These comparisons were based on laboratory jar tests and full scale plant operation using each coagulant. The purpose of these tests was to compare the effectiveness and cost of the two coagulants for trihalomethane (THM) reduction in the product water. In addition, the two coagulants were compared in their effect on dissolved organic carbon (DOC), total organic halogen (TOX), color, and aluminum residuals in the product water.

The typical sequence of jar test to plant scale test was followed but not on a seasonal basis due to schedule demands. The first jar test conducted for Olga was for the wet season beginning in the summer of 1984. When this was complete and the schedule for plant tests was coordinated with university and county personnel the dry season had begun. Consequently the bench scale experimentation was not seasonally ordered with plant scale investigation, however this did not hinder plant scale work since (as the data indicates) optimum conditions (pH and dose) for organic removal were similar in the wet and dry season. The coagulant and Cl_2 dose did vary by season as will be shown but operating conditions could be identified on site by considering raw DOC and color relative to previous bench work.

The decision was made to try to control the coagulation pH to 5.5 to 6.0 since the jar tests had shown this was the optimum pH range for organic removal. The coagulant was dosed at 80, 160, and 230-mg/L since these doses would cover varying range of organic removal as indicated by jar tests and experience. Most Florida coagulation plants use a alum dose and pH of approximately 100-mg/L and 5 to 7 for high color. These operating conditions provide a good basis for comparing PACl or alum coagulation of low turbidity and high color waters.

The Lee County Water Treatment Plant at Olga, Florida is located on the Caloosahatchee River approximately 10 miles east of Fort Meyers, Florida. The river drains a large watershed between Lake Okeechobee and the Gulf of Mexico as well as Lake Okeechobee itself. It is primarily agricultural and the site of the Florida sugar cane industry. Flow in the river in controlled by a series of four locks, used to control salinity and water level.

The actual wet season pH, total hardness, alkalinity, color and turbidity is shown in Figure 59. Note should be given to the type of water which is very typical of Florida surface waters. There is almost no turbidity in this water as turbidity varies from approximately 2 to 6 NTU's. The true turbidity may be even lower as these values represent plant readings and are not corrected for possible color interference. Such correction would require turbidity to be determined by difference before and after filtration. Color is above 100 cpu, adequate buffering is available (Alk > 90-mg/L $CaCO_3$), and moderate hardness (160 to 120-mg/L $CaCO_3$). Such waters do not require softening but do require color removal treatments typically accomplished by dosing alum such that the coagulation pH is between 5 and 7. Fortunately such treatment removes significant organics and minimizes Al residual. This water is a high color, low turbidity water which is treated by sweep coagulation. Floc sedimentation during sweep coagulation is not dependent on electrical destabilization of initial turbidity but enmeshment of organics in a massive metal hydroxide

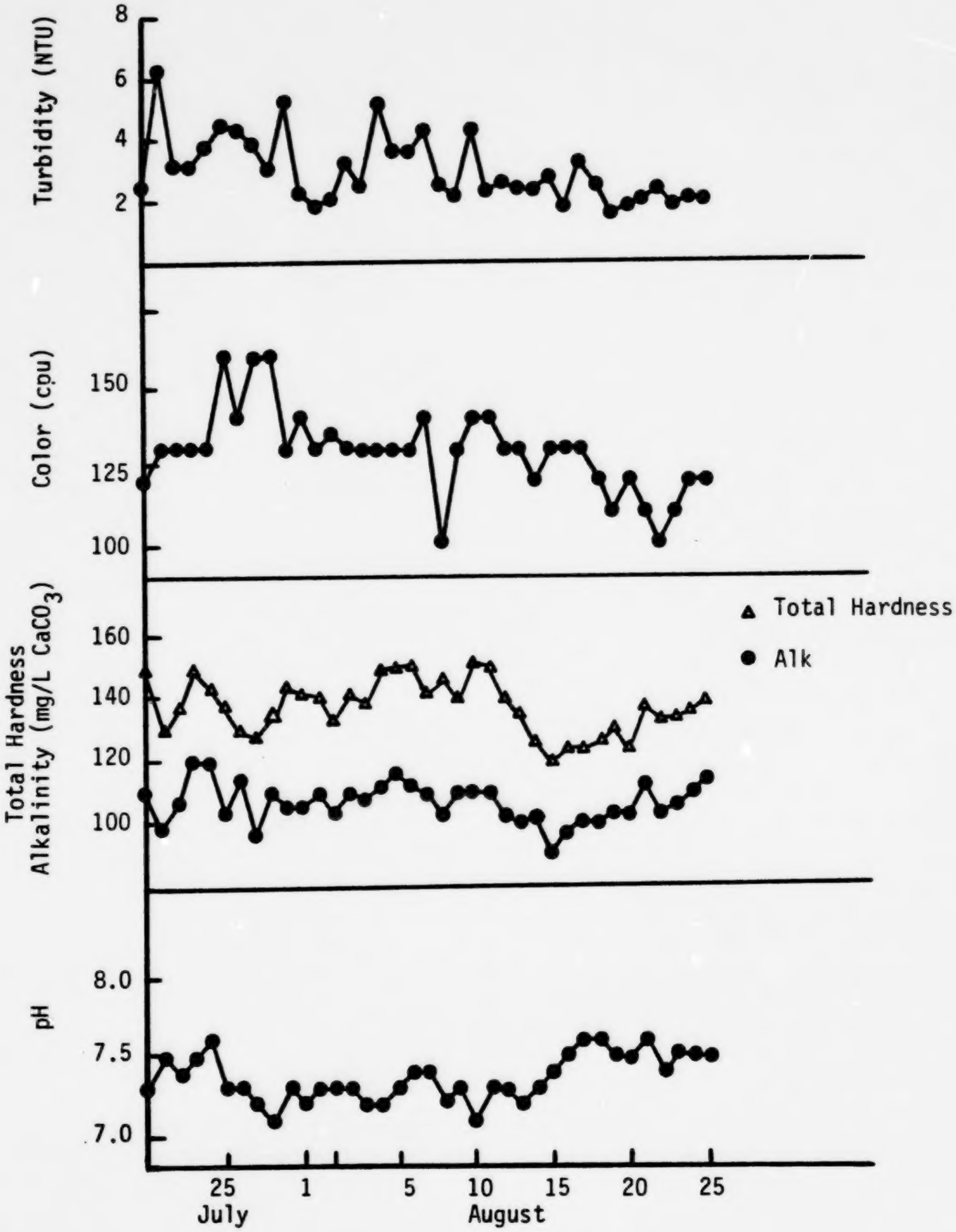


Figure 59 Raw water turbidity, color, alkalinity and pH during the wet season plant test.

floc. This provides a basis for the range of coagulant doses studied in plant testing.

A schematic layout of the plant and the points at which various chemicals were added is presented in Figure 60. The operating procedures of the plant previous to the time of both the wet and dry season plant tests included the following:

1. Alum addition at the raw water intake; ClO_2 has been added at the raw water intake as needed to control taste and odor problems during algal blooms. However, ClO_2 was not added during any plant test.
2. Lime and polymer addition in the first reactor at pH 10.3 for alum softening and limited odor removal and settling.
3. Aeration and massive Cl_2 addition for final color removal.
4. Settling in the second reactor.
5. Hexametaphosphate.
6. Filtration.
7. GAC filtration as needed.
8. Final chlorination.

During the dry season tests chemical feeds at the raw water inlet were halted except for the addition of 20-mg/L alum for the last four days of the dry season test to improve settling in the lime-softening unit. Either alum or PACl was added to the second reactor with sulfuric acid to lower the coagulation pH. Soda ash and hexametaphosphate were added to the effluent of reactor #2 prior to filtration. The purpose of the soda ash was to raise the pH to the calcium carbonate saturation level (pH_s) prior to filtration. Additional soda ash was added in the transfer basin to offset the pH drop caused by the addition of chlorine which was also added in the transfer basin. The major difference in the wet and dry season work was (1) more coagulant was needed in the wet season due to the increased organic load and (2) the addition of softening in a separate reactor from coagulation to reduce the high hardness in the dry season raw water. The typical plant operation did feed alum in the raw water but attempted to soften and coagulate in the same reactor at pH 10.3 which resulted in high Al residuals in the finished water. The dry season work essentially separated these two processes.

Full scale plant tests were conducted at the Olga Plant during the wet season from July 24 through August 25. The sequence of operations in reference to Figure 60 was:

1. Alum/PACl addition at the raw water intake along with acid as needed to reduce pH to approximately 5.5.
2. Polymer addition in reactor #2, flocculation and settling.
3. Aeration.
4. Lime addition to adjust pH to pH_s in accelator #2.
5. Soda ash and hexametaphosphate for stabilization.
6. Filtration through dual media filters.
7. Optional GAC filtration to control taste and odor (only on-line for 3 days of test).
8. Soda ash for final stabilization.

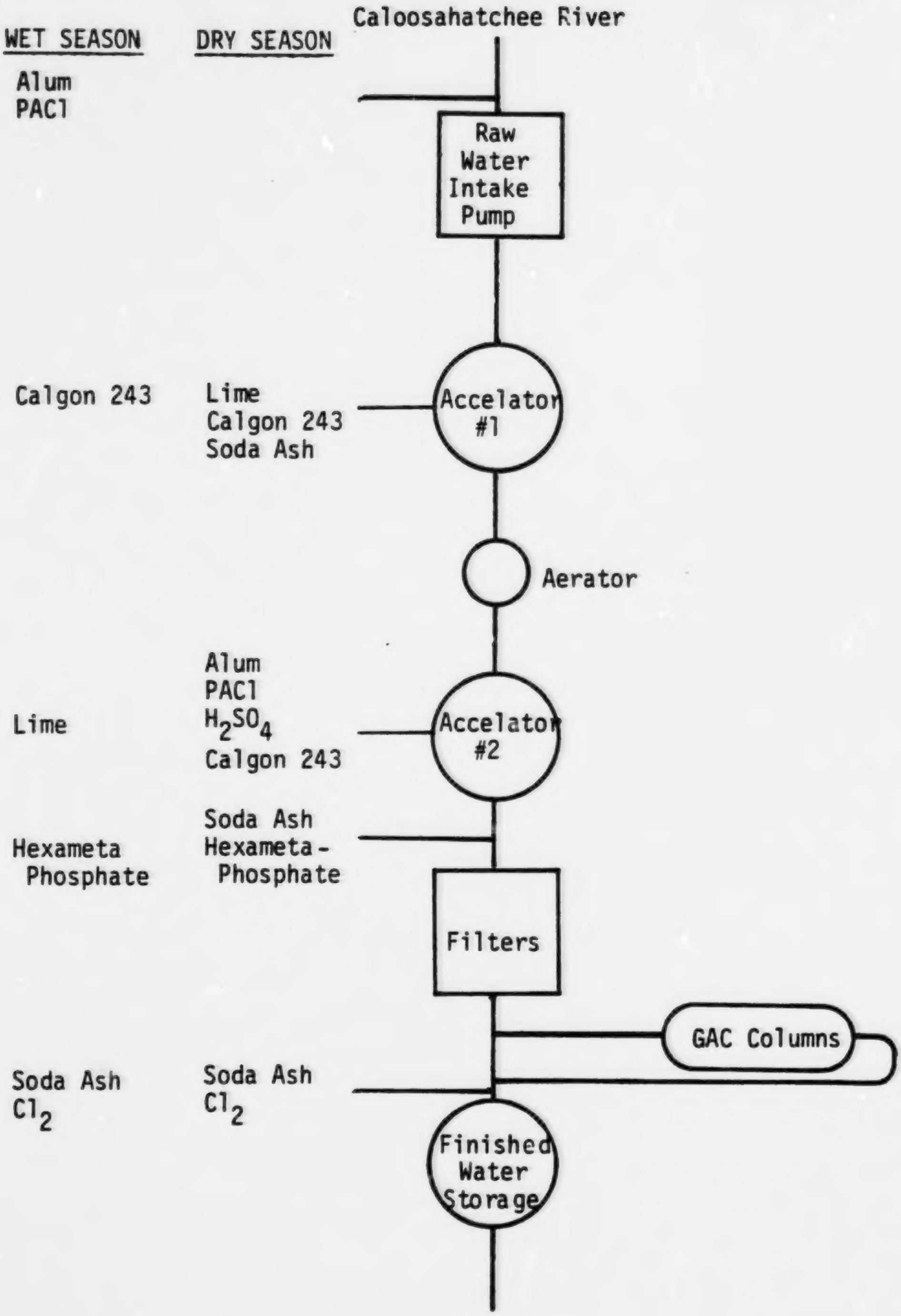

Caloosahatchee River
WET SEASON
DRY SEASON
Alum
PACl
Raw Water Intake Pump
Calgon 243
Lime
Calgon 243
Soda Ash
Accelator #1
Aerator
Lime
Alum
PACl
H_2SO_4
Calgon 243
Accelator #2
Hexameta Phosphate
Soda Ash
Hexameta-Phosphate
Filters
GAC Columns
Soda Ash
Cl_2
Soda Ash
Cl_2
Finished Water Storage

Figure 60 Olga water treatment chemical feed sequence by season and Plant Unit Operations.

9. Final chlorination.

Alum dose varied between 230 and 80-mg/L. After August 8, sulfuric acid was also available to lower the coagulation pH. The wet season coagulant doses were higher relative to dry season due to the increased organics. On July 26, reactor #1 was drawn down to clean out lime sludge which was slowly dissolving and interfering with pH control. The high alum dose on that day is a result of initial heavy feed rate to establish solids in the reactor.

6.2.7.2 Wet Season DOC

The wet season plant test at Olga ran from July 25 through August 25, 1984. DOC data for this period are presented in Table A-10 and Figure 61. DOC in the raw water varied between 17 and 25-mg/L during the study period. The average raw water DOC during the alum coagulation tests was slightly higher (2-mg/L) than the average raw water DOC for the PACl demonstration.

DOC reduction varied from 74% to 48% in the coagulation process and the DOC concentration in the lab tap sample followed closely the DOC in the coagulated and settled water. The initial DOC drop in the effluent DOC attributable to coagulation occurred on July 26, when the DOC dropped to less than 5-mg/L. This drop occurred after the reactor was drained and cleaned of any lime-softening sludge remaining from normal operation. The average plant DOCs for the different operating conditions are presented in Table 52. The significance of filtration in the plant is shown in Figure 61. The lower plant tap DOC relative to the reactor effluent is due to the filtration. The average decrease in filtered DOC was 1.0-mg/L and represented an average 12% drop in DOC. THM's were not measured on the reactor effluent but can be assumed to decrease by 10 to 15% due to filtration. At high coagulant doses, alum performed significantly better than the equivalent dose of PACl. For a dose of 160-mg/L with added sulfuric acid, the two coagulants performed equally well, however alum performed better than PACl when the acid feed was stopped (7.73 vs 9.36-mg/L settled DOC). The improved performance of alum over PACl when acid feed was discontinued is due to the lower pH achieved by alum (6.2-6.4) vs PACl (6.8). At lower coagulant doses (80 and 120-mg/L), alum was also more effective than PACl due to the lower pH reached with alum.

DOC concentrations in the plant effluent were similar to those in the wet season jar tests at similar coagulant doses and pH. The DOC reduction varied from in excess of 70% at 230-mg/L alum and pH 5.6 to 5.9 and slightly over 55% for 80-mg/L alum and pH 5.7. The pH variation for any alum dose was only from 5.6 to 6.4 for the wet season plant scale tests. Again, the intent was not to vary pH in the plant test as the bench tests had shown pHs of 5.5 to 6.0 were maximum for organic removal. In the pH range covered (5.9-6.4) there appears to be a general 8 to 10% DOC drop for every 80-mg/L alum dosed, beginning at 55% DOC removal at 80-mg/L, 65% removal at 160-mg/L, and almost 75% at 230. The plant data is variable but a definite increase in DOC removal with alum dose is apparent and does represent a definite THM precursor drop for increasing coagulant dose. For an

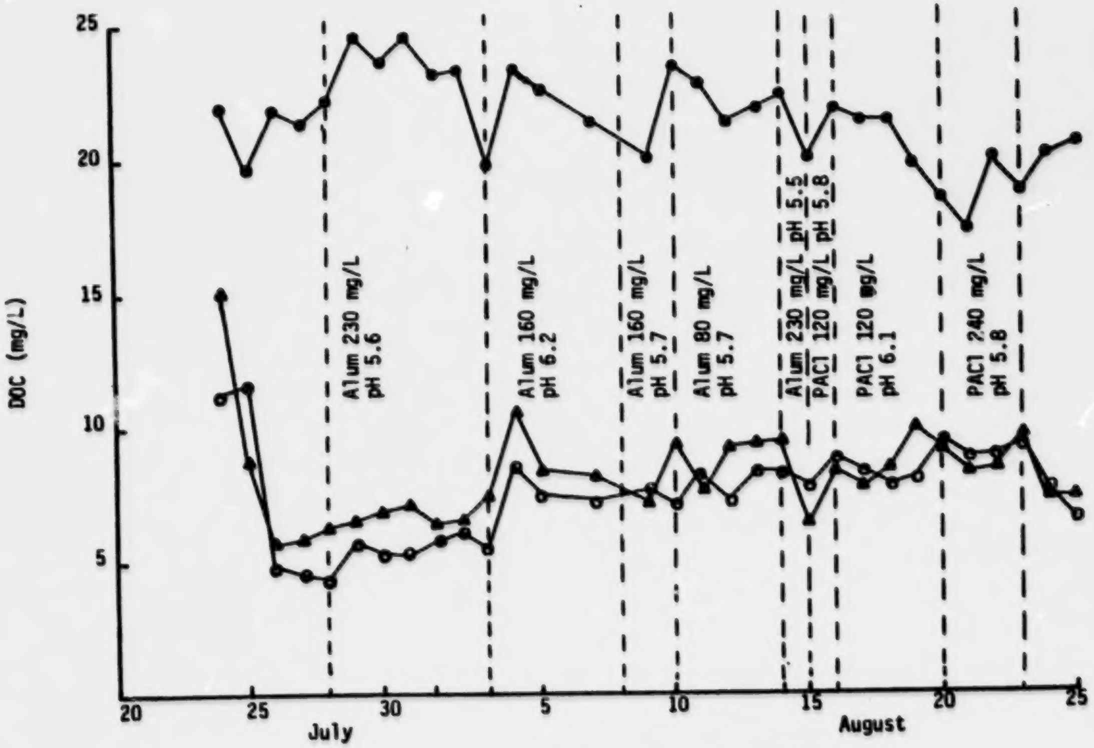


Figure 61 (●) Raw Water DOC (○) Accelator #1 DOC, and (△) Finished Water DOC at Olga Water Treatment Plant during Wet Season plant trials.

TABLE 52. Olga Plant Test Wet Season DOC Average and Standard Deviations (mg/L)

	Dose*	Acid	Average	Standard Deviation
Alum	230	-	2.33	0.18
	160	-	7.73	0.39
	160	+	7.63	0.29
	80	+	7.87	0.31
PACl	240	+	7.05	0.61
	160	+	7.93	0.12
	160	-	9.37	---
	120	+	8.96	0.18
	80	+	8.73	---

*mg/L $Al_2(SO_4)_3 .14H_2O$

additional cost of \$0.05/1000 gallons an additional 10% DOC reduction can be obtained which will increase THM precursor removal.

PACl is not quite as effective as alum for DOC removal. The PACl dose at 240-mg/L, pH 5.8 removed 68% of the initial DOC relative to 74% for alum. PACl doses of 160 and 120-mg/L removed only an average 56% and 53% of the DOC respectively. The 80-mg/L dose could not be sustained because the PACl floc would not settle. Floc settlability was achieved when the PACl dose was increased to 120-mg/L. A PACl increase of 80-mg/L from 160 to 240-mg/L did remove an additional 6% DOC.

There was some interest at Olga in determining the effect of the plant aerator on DOC reduction. The aerator preceded chlorination as chlorine was applied immediately after aeration in a large underground pipe going to rapid mix in the second reactor. Plant personnel described this point as the throat of the aerator and at one point project personnel thought chlorination may have preceded aeration and some organic reduction may have occurred. A total of 12 DOC measurements were made during late July and early August and showed no difference in DOC in and out of the aerator.

6.2.7.3 Wet Season THM

THM data for the summer plant demonstration at Olga are given in Table A-11 and Figure 62. Figure 62 gives the THM concentration at the Olga Plant tap for the period of the study. THM's declined dramatically (30 to 75%) during the first 2 weeks of the plant trials, particularly after the accelator #1 was cleaned on July 26th. The gradual decline in THM's was the result of flushing in the storage tank as well as the gradual decrease in chlorine dose combined with the change from a combined lime/alum process to an optimized coagulation process. Using the last 3 days of the high alum dose (230-mg/L), the average THM concentration at the tap was 194 μg/L. The average THM concentration for the equivalent PACl dose was 206 μg/L.

Using an alum dose of 160-mg/L, the THM concentration was 184 μg/L without acid feed and 165 μg/L with acid feed. Comparable concentrations for PACl were 329 μg/L and 207 μg/L, respectively. The lowest dose used with PACl was 120-mg/L which resulted in THM concentrations averaging 267 μg/L compared to 182 μg/L with alum. Given the expected variance in THM measurements the data does indicate alum did remove slightly more of the THM precursors than did PACl. THM's are affected by Cl_2 dose as well as precursors as shown in Figure 62. The expected THM precursor reduction is no more than 5% comparing alum to PACl.

The distribution system THM values in Table A-11 show a major drop in THM's once the plant was optimized for organic removal. Variable residence times make it impossible to determine the effect of PACl or alum on THM in the distribution system. The THM's at the plant tap ranged from 162 to 344 μg/L and from 229 to 450 μg/L in the distribution system. Cl_2 residuals were present in all but a few (<10) samples. This data indicates that

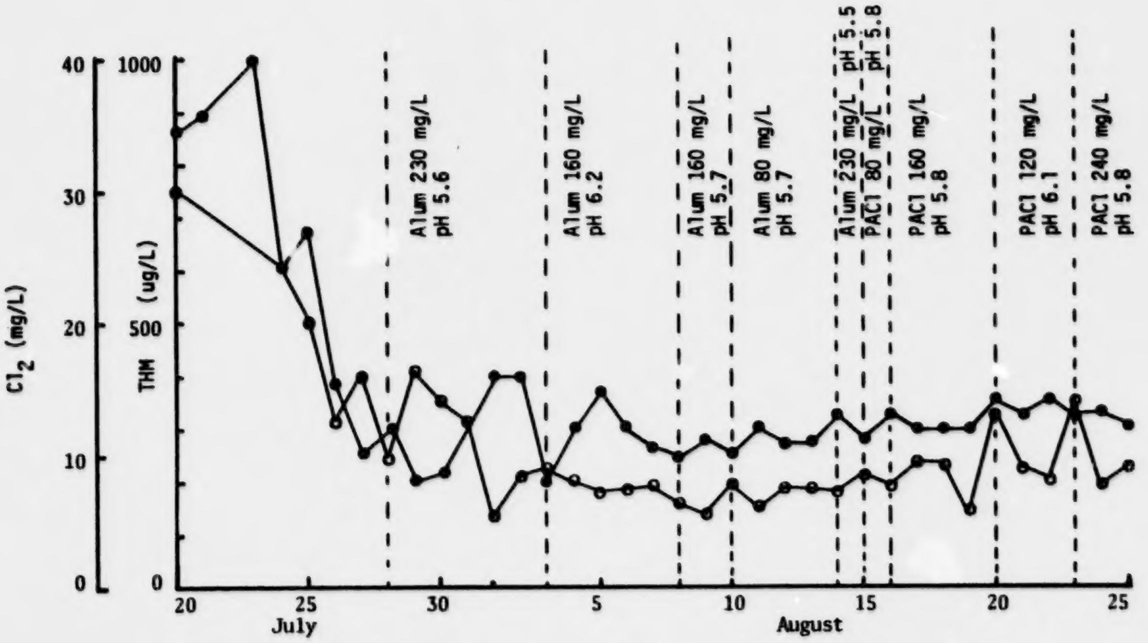


Figure 62 (•) THM Concentration and (•)Cl_2 dose at the Olga Water Treatment Plant during wet season plant trials.

2/3rds of the THM's are formed in the Olga Plant and 1/3rd in the distribution system. A smaller fraction of THM's would be formed at the plant if the plant's storage time (1 to 2 days) was less.

The Cl_2 dose was approximately constant by August 2. From that point the lowest THM's were recorded when alum was being used. THM's averaged 185 µg/L and ranged from 229 to 147 µg/L from August 2-15. When PACl was being used this range was 351-148 µg/L and the average was 235. This is a 21% decrease on a plant basis for THM reduction which was not seen in the jar tests. However, both the jar test and the plant test in the wet season indicate alum removes more THM precursors than PACl.

6.2.7.4 Wet Season Plant TOX

TOX values for the wet season plant demonstration are shown in Figure 63. The average TOX value for 230-mg/L alum dose was 737 µg/L and for an equivalent PACl dose was 721 µg/L. Other than the sample taken on July 24, the day before alum coagulation started, no significant difference in the TOX in the finished water resulting from alum or PACl coagulation.

The ratio of TOX to DOC and THM's and the ratio of TOX to THM is presented for the wet season PACl alum tests in Table 53. The average DOC:THM:TOX ratio is 1:34:92 and there was significant variation (1:48:110 to 1:15:63). The ratios are based on DOC in mg/L C, TOX in µg/L Cl, and THM's in µg/L. The THM values are from samples taken at the plant tap as were the DOC and TOX values. The average THM:TOX ratio was 1:2.9 and may be more significant because both of these parameters increase directly with Cl_2 dose where DOC does not. For the wet season this data indicates the ratio of THM:TOX was about 1:3 and the ratio of DOC:THM:TOX was about 1:30:90.

6.2.7.5 Wet Season Color

Color data for the summer plant tests are shown in Figure 64. Raw water color declined from as high as 170 cpu at the beginning of the study to 105 cpu at the end. In all cases, the finished water at the plant tap had a color concentration of less than 5 cpu. Settled color from coagulation averaged 8.5 cpu at an alum dose of 230-mg/L with no acid feed vs. 6 cpu for an equivalent PACl dose with acid feed. At an alum dose of 160-mg/L with no acid feed the color averaged 18 cpu and with acid feed the color averaged less than 2 cpu. The equivalent conditions for the PACl values were 11 and 15 cpu, respectively. At an alum dose of 80-mg/L with acid feed, the resulting color averaged 10 cpu and for a 120-mg/L dose of PACl the color averaged 13 cpu. Consequently, pH was more significant than dose for color removal.

The plant chlorination process did reduce color. Once the Cl_2 dose was stabilized it averaged 12 to 14-mg/L during August testing. There was approximately 0.5 cpu of color destroyed for every-mg/L of Cl_2 added during this period. It is easy to envision Cl_2 doses of 40-mg/L, as was the practice, being necessary to reduce settled colors of 30 or more to below the 15 cpu regulation.

TABLE 53. DOC, THM and THM Ratios for Wet Season

DOC mg/L	THM ug/L	TOX ug/L	DOC/THM/TOX	TOX/THM
15.1	605	1173	1/40/78	1.9
5.7	395	740	1/69/130	1.9
6.3	250	634	1/40/101	2.5
6.5	514	712	1/64/110	1.7
7.1	340	782	1/48/110	2.3
9.4	187	561	1/20/60	2.0
9.6	180	776	1/19/81	4.3
6.5	229	891	1/35/137	3.9
8.3	195	578	1/23/70	3.0
7.9	234	838	1/30/106	3.6
8.6	238	705	1/28/82	3.0
10.0	148	626	1/15/63	2.2
9.2	329	574	1/36/62	1.7
8.4	236	638	1/28/76	2.7
8.5	209	842	1/25/99	4.0
9.8	351	917	1/36/94	2.6
7.5	187	766	1/25/107	4.1
7.3	225	675	1/31/92	3.0
Average			1/34/92.1	2.9

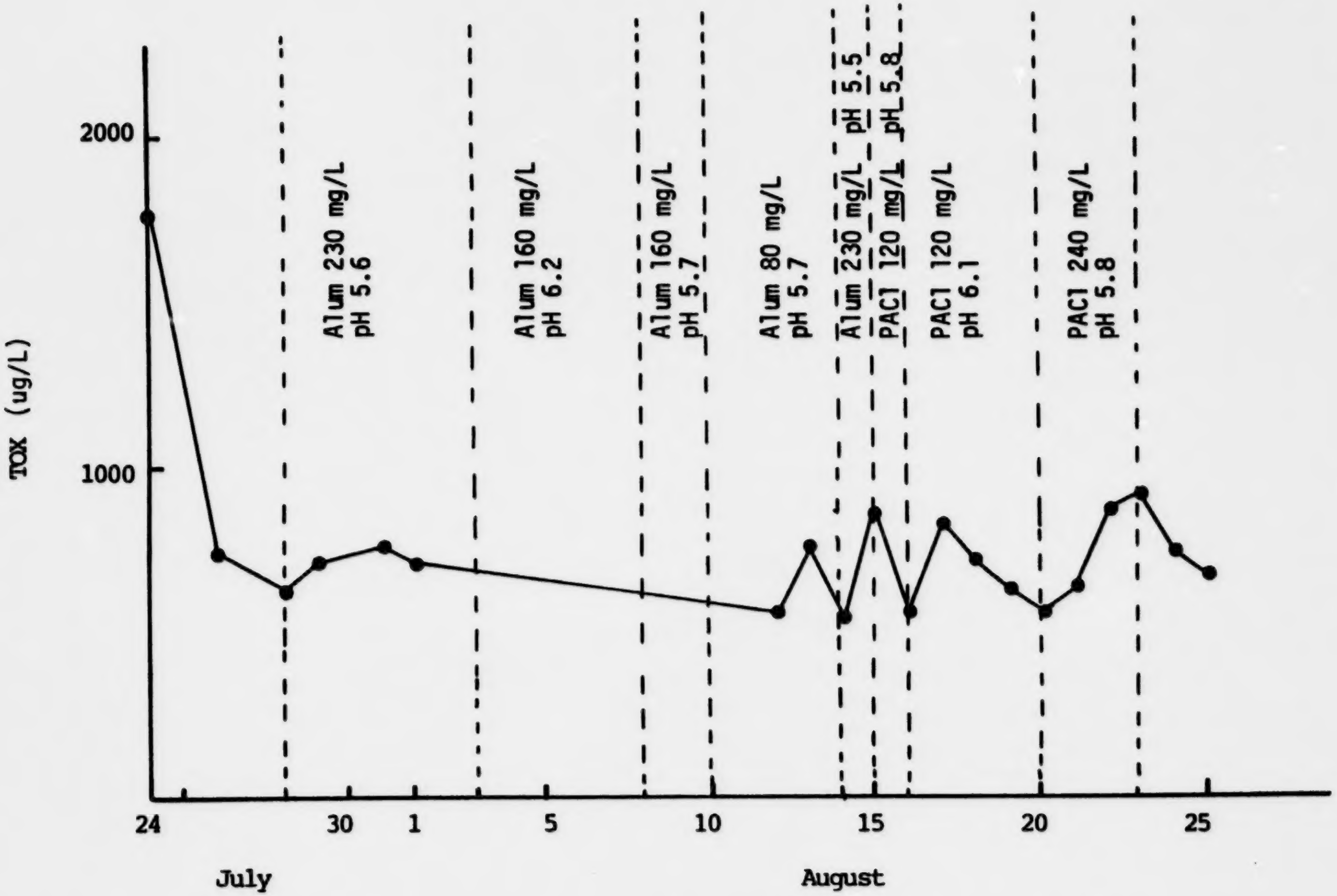


Figure 63 TOX concentrations in Olga Finished Water during wet season.

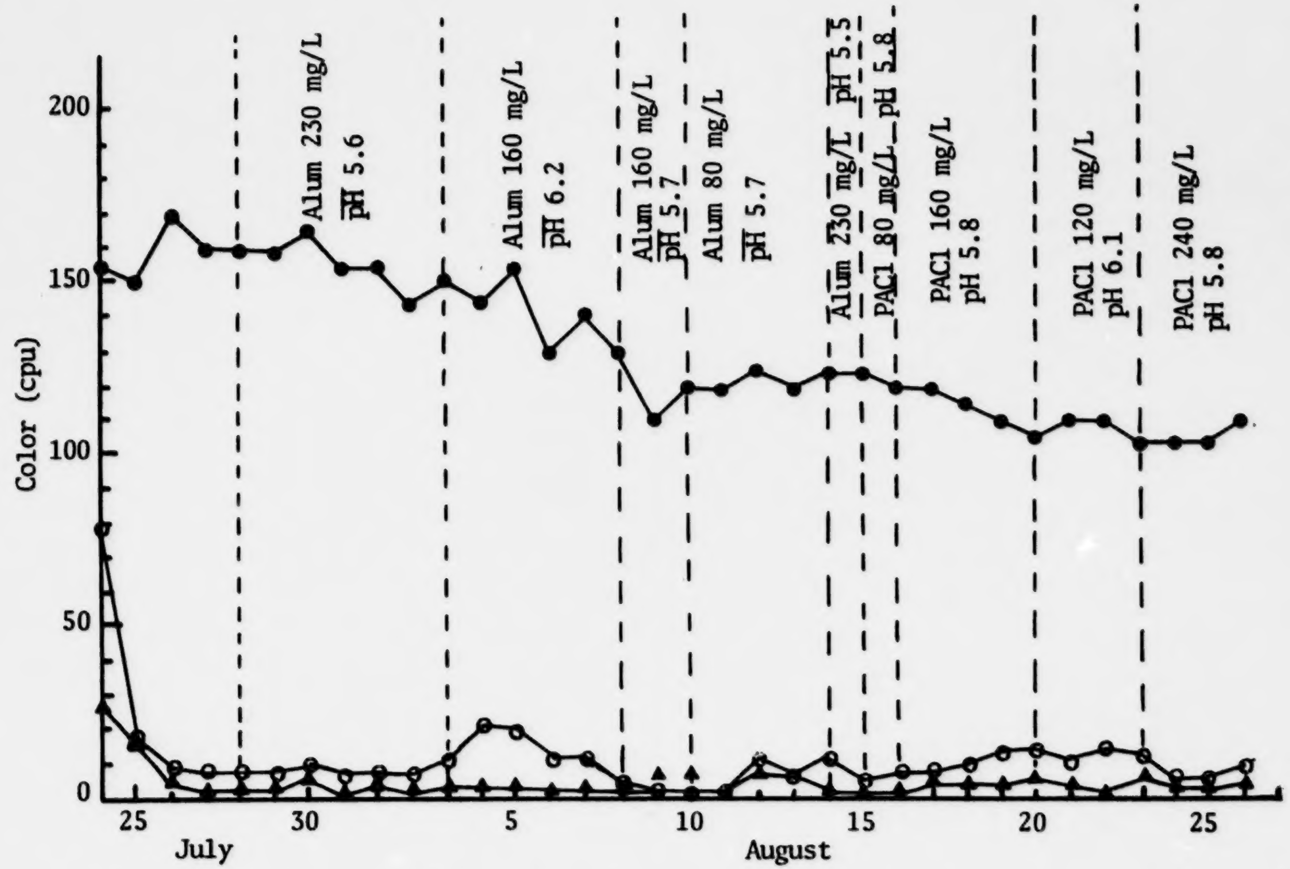


Figure 64 Raw (●), Settled (○) and TAP (▲) Color for Olga Water Plant during the wet season.

Color removal from the reactor averaged 94%, 93%, and 96% for an alum doses of 80-mg/L, 160-mg/L and 230-mg/L respectively. Two observations can be made. First, color removal was essentially complete at the lowest dose utilized and second, color is not a good surrogate for THM, TOX, or DOC. This is reflective of jar test data which had shown that alum doses over the same range removed from 93% to 98% of the color. For the wet season PACl/alum testing color would not be a satisfactory indicator of DOC, THM or TOX precursor removal because increasing coagulant dose increased organic removal and had no significant effect on color removal. Color should not be used by plant personnel as a surrogate for THM precursor removal. PACl doses of 120, 160, and 240-mg/L average 89%, 89%, and 96% color removal, respectively. There is slightly less color removal when compared to alum but the same observation of not using the color as an indirect measurement of DOC, TOX, or THM precursors is valid.

6.2.7.6 Other Wet Season Parameters

Aluminum residuals for the lab tap are given in Figure 65. The initial value for aluminum residual during lime softening was 1.43-mg/L. After the #1 accelerator was cleaned on July 26, this value decreased to 0.3-mg/L or less for most of the study. On August 16, when the plant operation was switched to PACl, the plant concentration rose to 0.41-mg/L as a result of floc carry over at the 80-mg/L PACl dose. The average aluminum concentration for alum was 0.19-mg/L while the average aluminum residual using PACl as a coagulant was 0.20-mg/L. Aluminum residual in the raw water averaged approximately 0.20-mg/L. The average values obtained for aluminum residual compared well with those found during the wet season jar tests under similar conditions. The data very clearly show there is no significant difference between Al residuals after PACl or alum coagulation. The plant data as well as the jar test data show very clearly the Al residuals after coagulation were controlled by coagulation pH and were minimized at pH 5 to 7. Such operating conditions will ensure no consumer complaints due to high Al residuals in the finished water.

During the wet season tests the plant was operated approximately 20 hrs/day. Plant shutdown was controlled by water storage and demand by time of day in the plant reservoir and was commenced to keep finished water from being wasted. The flow during operation was between 1600 and 1800 gpm which meant that surface loading rates varied between 0.43 and 0.48 gpm/ft^2 for the settling area. The surface loading ratio on the filters were from 1.8 to 2.0 gpm/ft^2. Since normal design rates for settling and filtration are 0.5 gpm/ft^2 and 2 gpm/ft^2 no hydraulic stress was placed on plant performance during testing.

The turbidities from each reactor and the finished turbidities are shown in Table 54. The PACl and alum coagulation occurred in reactor #1. The turbidity data from this reactor indicates that very low turbidities were obtained from reactor #1, (typically ≤ 1 NTU) and there were no differences in the turbidity carry over due to either PACl or alum coagulation once a settleable floc was formed. However, Alum formed a settleable floc at a lower dose than PACl. There was a periodic problem with small clouds of alum or PACl floc often late in the day, however the carry over

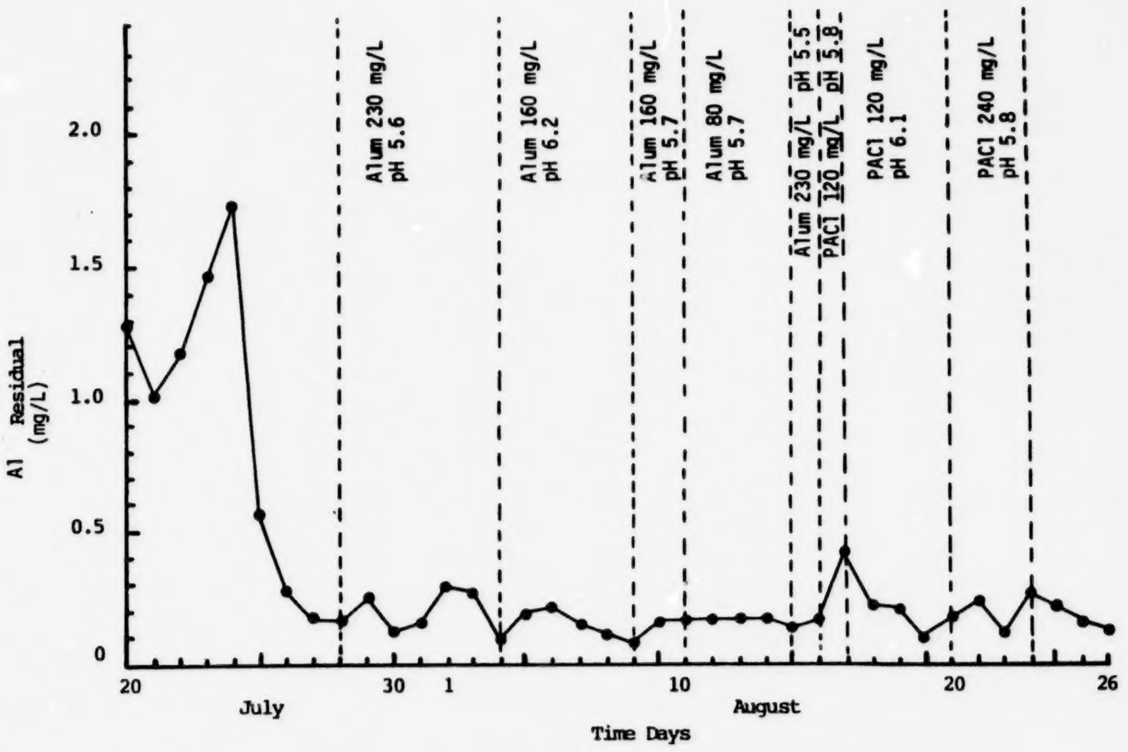


Figure 65 Aluminum Residual in Olga Finished Water during Wet Season

TABLE 54. Wet Season Turbidity (ntu)

Date	Coagulation Reactor	Settling Reactor	Finished
7/24*	5.1	5.0	0.8
7/25	1.1	5.0	1.4
7/26	1.8	11.0	1.0
7/27	1.8	3.7	0.8
7/28	0.6	1.9	0.8
7/29	0.4	0.9	0.2
7/30	0.3	0.6	0.3
7/31	0.5	0.8	0.3
8/1	0.5	1.5	0.2
8/2	0.3	1.2	0.2
8/3	0.4	2.9	0.1
8/4	0.6	5.2	0.1
8/5	0.6	4.2	0.3
8/6	0.5	3.9	0.4
8/7	0.5	5.4	0.4
8/8	0.4	2.9	0.4
8/9	1.0	2.4	0.3
8/10	0.6	1.5	0.2
8/11	1.0	2.2	0.1
8/12	0.4	1.7	0.1
8/13	0.6	1.2	0.2
8/14	0.8	0.9	0.2
8/15	0.8	1.4	0.2
8/16**	1.5	1.8	0.1
8/17	0.5	1.4	0.3
8/18	0.4	1.8	0.2
8/19	0.3	1.6	0.2
8/20	0.5	4.2	0.3
8/21	0.6	2.2	0.4
8/22	0.7	1.0	0.2
8/23	0.7	2.7	0.3
8/24	0.7	1.0	0.3
8/25	0.2	0.7	0.1

* Alum coagulation to 7/24

** PACl coagulation to 8/25

from this snowflake floc did not effect turbidity measurements. The highest turbidity was from reactor #2 which received only CaO to stabilize pH. Small $CaCO_3$ floc would form in the reactor due to either poor mixing at the point of CaO feed or excessive CaO feed in an attempt to offset the pH drop due to chlorination. The later would not work until all HCO_3^- buffering was exceeded and in normal practice often led to pHs of 10 or more onto the filters. Such practice turned the filters into a softening process. This is the reason the Olga Plant filters are frequently rebuilt.

The sludge volume was determined daily for all plant tests and is presented for the wet season in Table 55. The coagulant (PACl or alum), dose, sludge volume and dose to volume ratio are given in Table 55. The percent of sludge remaining after 10 minutes of settling in a standard one liter sludge sample is the sludge volume. The samples were always taken from the center of the reactor in the rapid mix section by plant personnel.

The data is quite variable but does show a large increase in sludge volume once plant testing was commenced. This would be expected as an alum sludge replaced a lime sludge from July 24 to the 27. These dates correspond to a sludge volume increase of 4% to 46% (10x). The volume of the sludge samples is affected by coagulant dose and sludge discharges from the reactor. However these operating conditions were relatively similar for periods of testing PACl and alum in the wet season. The ratios of coagulant dose/volume vary from 48 to 2 for alum averaging 11, and vary from 30 to 3 for PACl averaging 12. These data indicate there was no significant difference in the sludge volume produced by either alum or PACl and that problems of alum sludge disposal would be very similar to disposal of PACl sludge.

6.2.7.7 Wet Season Summary

The DOC and color removal during plant tests for the wet season were similar to jar tests results even though the wet season jar and plant data were developed from different periods of the wet season. The coagulation pH varied from 6.8 to 5.6 and the coagulants were dosed at 239, 160 and 80-mg/L in as $Al_2(SO_4)_3 \cdot 14\ H_2O$. An alum dose of 230-mg/L at a pH of 5.9 to 5.6 averaged over 70% DOC removal and 95% color removal. The isopleths indicated a color removal of 93% or more and DOC removal in excess of 70% for these conditions. At the lower plant alum dose of 80-mg/L at pH 5.7 which required acid feed, the typical color and DOC removal was 94% and 57% which is very close to the isopleth DOC removal and about 5% better than the isopleth color removal. The PACl plant results were also close to the PACl isopleths. PACl doses of 160 and 240-mg/L at pH 5.8 average near 58% and 63% DOC removal and about 90% and 96% color removal. Jar test data indicate PACl should remove 65 to 70% of the raw DOC and 93 to 96% of the raw color.

The purpose of varying the coagulant type, dose, and pH was to determine in a plant scale the effect of those parameters on color, DOC and THM precursor removal. The results clearly show the same effects in jars as in the plant. The best DOC removals (70+%) corresponded to the highest alum dose (230-mg/L) and a low coagulation pH (5.7). PACl doses of 230-mg/L

TABLE 55. Settled Sludge Volume for Wet Season Olga Plant Tests

Sludge Solids

Date		Coagulant	Dose* mg/L	Sludge Volume %	Dose/Volume%
July 24,	1985	Alum	190	4	48
" 25,	"	"	230	12	19
" 26,	"	"	320	35	9
" 27,	"	"	165	46	4
" 28,	"	"	190	40	5
" 29,	"	"	230	40	6
" 30,	"	"	230	35	7
" 31,	"	"	230	50	5
Aug. 1,	1985	"	230	41	6
" 2,	"	"	230	34	7
" 3,	"	"	230	30	8
" 4,	"	"	160	14	11
" 5,	"	"	160	8	20
" 6,	"	"	160	20	8
" 7,	"	"	160	23	8
" 8,	"	"	160	28	6
" 9,	"	"	160	11	15
" 10,	"	"	160	9	21
" 11,	"	"	160	17	9
" 12,	"	"	80	38	2
" 13,	"	"	80	16	5
" 14,	"	"	80	7	11
" 15,	"	"	230	8	29
" 16,	"	PACl	80	7	11
" 17,	"	"	160	8	20
" 18,	"	"	160	23	8
" 19,	"	"	160	38	4
" 20,	"	"	160	38	4
" 21,	"	"	120	11	11
" 22,	"	"	120	7	17
" 23,	"	"	120	5	24
" 24,	"	"	240	8	30
" 25,	"	"	240	40	6
" 26,	"	Alum	160	48	3
" 27,	"	"	160	42	4

removed slightly more than 60% of the DOC at pH 5.8. PACl doses of 160-mg/L at pH 5.8 average 58% DOC removal whereas alum removed about 5% more DOC at the same conditions. The wet season plant and jar test data indicate that 50 to 70% of the DOC can be removed during coagulation, and that coagulation can be maximized for organic removal, however, not to the extent that the THM MCL can be met for this water.

THM precursor reduction was more complex that either color or DOC removal because of the varying plant Cl_2 dose and the natural delay after chlorination caused by plant storage. The flow through the plant reservoirs is not predictable by either a plug flow or complete mix model so the THM's are difficult to correlate to coagulant dose and pH. However the THM data at the plant tap varied from about 350 to 150 μg/L after the study commenced. This represents a THM decrease relative to the normal plant THM's of 50 to 75%. The THM data from the plant wet season work show that alum or PACl is effective for precursor removal when dose and pH are controlled. The data clearly show that neither alum nor PACl can be used in the wet season to meet the 0.1-mg/L MCL and the PACl offers no advantage over alum under these conditions.

TOX precursor removal varied directly with the removal of other organic parameters (THM precursor, DOC, color) during coagulation. The average ratio of THM to TOX was approximately 1 to 3 and there was ±1 part TOX variation. The point should be noted that even though all the removal of the organic parameters (color, DOC, THM and TOX precursors) were directly related, color removal was essentially complete at a much lower coagulant dose than DOC, THM or TOX precursor removal. Therefore color should not be used as a surrogate for THM precursors. DOC could be used as a surrogate for THM or TOX precursor by plant personnel.

Al residuals were controlled by coagulation pH to less than 0.5-mg/L for either alum or PACl. In summary this data could indicate that alum may have a slight advantage (+5%) over PACl for color, DOC, THM or TOX precursor removal. However for practical significance in the treatment of these high organic low turbidity waters there is no difference.

6.2.8 Cost Summaries

Tables 56 and 57 give summaries of costs for the wet and dry season plant demonstrations at Olga. The procedure used for calculation based the cost of each item on the amount of finished water actually pumped from the plant rather than the amount of raw water pumped.

The total power and labor costs for July/August (wet season) and February/March (dry season) were taken from plant records and the total product water pumped in each period. This gave power and labor costs per thousand gallons. The total mass of lime, chlorine, alum, PACl, RS-2 (stabilizer) and CA-243 (polymer) used each day was divided by the finished water produced and multiplied by the cost of each chemical. Daily costs were averaged for each treatment condition to give the costs per thousand gallons for each condition. GAC rental costs were calculated by dividing

TABLE 56. Wet Season Cost ($/1000 Gallons) Summary

Chemicals	$/lb	Standard Treatment	Alum			PACl		
			(80 mg/L)	(160 mg/L)	(230 mg/L)	(120 mg/L)	(160 mg/L)	(230 mg/L)
Lime	0.0340	0.0456	0.0036	0.0052	0.0046	0.0028	0.0044	0.0017
Chlorine	0.1300	0.0407	0.0138	0.0145	0.0141	0.0184	0.0148	0.0123
Alum	0.0740	0.0443	0.0598	0.1156	0.1508	--	--	--
PACl	0.2904	--	--	--	--	0.3122	0.4097	0.5817
RS-2	0.8000	0.0136	0.0164	0.0133	0.0131	0.0096	0.0151	0.0095
CA243	2.7100	0.0072	0.0126	0.0127	0.0100	0.0131	0.0122	0.0100
Soda Ash	0.1225	--	0.0513	0.0138	--	0.0743	0.0501	0.0611
Acid	0.0850	--	0.0285	0.0078	--	0.0322	0.0326	0.0303
Subtotal Chemical		0.1514	0.1860	0.1829	0.1526	0.4626	0.5389	0.7066
Labor		0.2692	0.2692	0.2692	0.2692	0.2692	0.2692	0.2692
Power		0.3105	0.3015	0.3015	0.3015	0.3015	0.3105	0.3015
Subtotal		0.7221	0.7567	0.7576	0.7633	1.0333	1.1096	1.2773
Rent on GAC		0.3483	0.3483	0.3483	0.3483	0.3483	0.3483	0.3483
Total		1.0704	1.1050	1.1019	1.1116	1.3816	1.4579	1.6256
Total Adjusted for Soda Ash and Acid				1.1381	1.1616			
Cost % Increase		--	3.2	6.3	8.5	29.7	36.2	51.8
THM (ug/L)		710	199	183	197	251	243	208
THM % Decrease		--	72	74	72	65	66	72

TABLE 57. Dry Season Cost ($/1000 Gallons) Summary

Chemicals	$/lb	Standard Treatment	Alum			PACl		
			(80 mg/L)	(100 mg/L)	(120 mg/L)	(80 mg/L)	(100 mg/L)	(120 mg/L)
Lime	0.0340	0.0459	0.0335	0.0338	0.0332	0.0338	0.0298	0.0291
Chlorine	0.1300	0.0170	0.0104	0.0108	0.0089	0.0106	0.0118	0.0111
Alum	0.0740	0.0427	0.0545	0.0669	0.0813	--	--	--
PACl	0.2904	--	--	--	--	0.2213	0.2830	0.3106
RS-2	0.8000	0.0115	0.0150	0.0082	0.0152	0.0130	0.0103	0.0127
CA243	2.7100	0.0066	0.0165	0.0173	0.0157	0.0154	0.0166	0.0156
Soda Ash	0.1225	--	0.12275	0.1225	0.1225	0.1225	0.1225	0.1225
Acid	0.0850	--	0.0337	0.0337	0.0337	0.0506	0.0506	0.0506
Subtotal		0.1237	0.2861	0.2932	0.3110	0.4672	0.5246	0.5522
Labor		0.1205	0.1205	0.1205	0.1205	0.1205	0.1205	0.1205
Power		0.1551	0.1551	0.1551	0.1551	0.1551	0.1551	0.1551
Subtotal		0.3993	0.5617	0.5688	0.5866	0.7428	0.8002	0.8278
Rent on GAC		0.1959	0.1959	0.1959	0.1959	0.1959	0.1959	0.1959
Total		0.5952	0.7576	0.7647	0.7825	0.9397	0.9961	1.0237
Cost % Increase			27.28	28.48	31.47	57.38	67.36	71.99
THM (ug/L)		286	195	215	173	208	201	203
THM % Decrease		--	32	25	40	27	30	29

the rental charge over each two month period by the total volume of finished water pumped.

In the wet season, acid and soda ash use started after the plant demonstration had begun. Therefore the cost of these chemicals are not included in the first chemical cost subtotaled in Table 56. The second subtotal includes the required cost of these chemicals to stabilize the water and regulate to pH. This second subtotal was used for cost comparison which includes stabilization as well as optimum organic removal.

In the wet season demonstration, chemical costs accounted for 14 to 43 percent of the total treatment cost. The wide variation was due to the cost difference between PACl and alum. Using alum increased the total treatment cost a maximum of 6 percent ($1.07 to $1.16) over the standard treatment. Using PACl increased the total cost from 29 to 52 percent more than the standard treatment.

Based on a plant THM average of 710 μg/L, the average percent THM reduction varied from 59% for alum at 230-mg/L to 74% for alum at 160-mg/L. The percent THM reduction varied between 65 and 71 percent when PACl was used. The actual reduction achieved with 230-mg/L alum is somewhat higher than indicated. This condition was the first tested during the wet season plant demonstration and therefore some THM's from the previous treatment existed in the system. Another consideration was the gradual lowering of chlorine dose to ensure Lee County officials that adequate disinfection would be maintained. Using data from the last three days of operation at this dose, the percent THM reduction was increased to 72 percent with the average THM concentration equal to 197 μg/L. This is shown in Table 58 and is more reflective of expected THM's following alum doses of 230-mg/L.

These data indicate that no advantage is obtained using alum doses greater than 80-mg/L provided adequate control of pH is maintained. Controlling pH using sulfuric acid is less expensive than increasing alum dose even when soda ash is used for final pH adjustment. PACl performed equally or slightly worse than alum under equivalent conditions. However, PACl use resulted in much higher costs than alum under all conditions.

Cost savings were achieved in reduced lime and chlorine use. Softening was not conducted during the wet season, because the raw water was relatively soft during this period. Chlorine use was reduced by over 60 percent by eliminating the need for high chlorine doses to remove color.

During the dry season plant tests a similar procedure was used to calculate chemical costs. Several differences should be noted between the wet and dry seasons. Power, labor, and GAC rental costs per 1000 gallons were almost 50% higher during the wet season than the dry season, although the monthly cost for these items was relatively constant for both seasons. The amount of finished water pumped increased in the dry season thus lowering the unit cost. Because of this, changing chemical costs had a greater effect on overall costs during the dry season than during the wet season. In addition, lime softening was conducted during the dry season and approximately 50 percent of the soda ash was used for lime softening

TABLE 58. PACl/Alum Summary

	PACl - Wet		PACl - Dry		Alum - Wet		Alum - Dry	
	%R	C	%R	C	%R	C	%R	C
DOC	57	8.6 mg/L	39	12.1 mg/L	58	7.0 mg/L	48	10.1 mg/L
THM	--	235 ug/L	--	204 ug/L	--	248 ug/L	--	173 ug/L
Color	91	11 cpu	78	12 cpu	94	9 cpu	84	10 cpu
THM	--	716 ug/L	--	380 ug/L	--	710 ug/L	--	421 ug/L
Cost $/1000 gal		1.07 - 1.62		0.60 - 1.02		1.07 - 1.16		0.60 - 0.78
% THM Reduced		71%		40%		72%		20%

Season	Wet	Dry
Ratios		
TOX/THM/TOX	1/34/92	1/16/39
THM/TOX	1/2.9	1/2.4

and 50 percent for stabilization. Cost increases varied between 27 and 31 percent for alum treatment and between 57 and 72 percent for PACl.

Overall, cost increases over normal treatment were higher and THM reduction lower during the dry season compared to the wet season. The lower percent reduction in THM formation was a result of lower THM's in the water produced by the normal treatment (286 µg/L vs 710 µg/L). It is possible that THM reductions could have been compared in the dry season if pH could have been controlled better during coagulation. While lime and soda ash cost increased in the dry season over the wet season, chlorine cost decreased due to the lower temperature and slower dissipation rates.

It should be noted that considerable cost savings at this plant could be achieved by eliminating the use of the GAC columns. Intermittent taste and odor problems could be controlled as they occur using powdered activated carbon with proper mixing and reaction time.

6.2.9 PACl/Alum

The PACl/alum investigations were made on a bench scale and on a plant scale at the Olga water treatment plant located in Lee County, Florida, for both the wet and dry seasons. The organic parameters that were investigated were DOC, THM, TOX, and color as a result of coagulant dose and pH. Aluminum residuals, turbidity, sludge volumes and dry season softening were also investigated.

The raw water color averaged about 57 cpu during the dry season tests and increased 228% to 130 cpu during the wet season tests. Raw water DOC averaged 19.6-mg/L during the dry season plant tests and increased 9% to 21.4-mg/L during the wet season test. The raw turbidity average 6 NTU during the wet season and 4 NTU during the dry season. The alkalinity and hardness during the wet season were approximately 140 and 110-mg/L as $CaCO_3$, and increased to 250 and 160-mg/L as $CaCO_3$ during the dry season. The raw water used during both plant tests was a low turbidity, high color water that would be normally treated by sweep coagulation for color removal only. Two PACl coagulants, 100S- a sulfate chloride base and 190- a chloride base coagulant, plus alum were compared for organic removal.

Jar testing during the wet and dry seasons demonstrated the optimum coagulation pH was between 5.0 and 6.0 for DOC and color removal for all coagulants. PACl 100S was superior to PACl 190 for the removal of DOC and color in jar tests and was selected for plant testing. DOC removal increased with increasing coagulant dose between pH 5.0 and 6.0 but appeared to approach a limiting value. Coagulant doses of 117-mg/L as alum removed 60% of the DOC and 90% of the color in the wet and dry season jar test. Increasing coagulant dose to 175-mg/L increased DOC removal 5% to 10% and had a lesser effect on color removal. The percent color and DOC removal during the wet season were slightly higher (<5%) than during the dry season for all coagulants. However the actual DOC concentration and, to a lesser extent, the color remaining in each season were approximately the same at equivalent conditions for each coagulant. Wet season THM's reduction was very similar to DOC removal. The optimum pH range for THM's reduction was 5.0 to 6.0 for all coagulants and increasing coagulant dose increased THM's reduction in all jar tests. Doses of 117 to 175-mg/L as $Al_2(SO_4)_3 \cdot 14H_2O$ realized 90% of the maximum THM's reduction during coagulation.

During the jar tests, alum generally removed slightly more (<5%) DOC, color, and THM's than PACl. The seasonal jar tests did not indicate a major difference in coagulant demand for organic removal. This may have been due to the small DOC increase during the wet season. Aluminum residuals after coagulation with either PACl-100S or alum were typically less than 0.4-mg/L Al and were controlled by coagulation pH. These levels would not cause post precipitation problems. Removal of DOC and THM's during coagulation should be in excess of 60% accompanied by more than 90% color removal. Although DOC, THM's and color removal vary directly with equivalent coagulation dose and pH, DOC is the better indicator of THM's than color.

Plant tests using alum and PACl were conducted in the dry and wet season for approximately 30 days each. The aluminum residuals in the distribution system were generally less than 0.2-mg/L. The aluminum residuals in the distribution system under normal plant process of using alum in conjunction with softening resulted in residuals of 5-mg/L Al. The seasonal plant data demonstrated finished Al residuals are controlled by coagulation pH which should be between 5.5 to 6.5 for minimum Al residuals. There was no significant difference between the sludge volumes produced during alum or PACl coagulation during the plant tests for equal operating conditions. The turbidity carryover from the settled water using either alum or PACl coagulation was typically less than 2 NTU and exhibited no differences between the two coagulants. A PACl 100S dose of 80-mg/L as alum was not sufficient to produce a settleable floc during the plant test. An equal alum dose produced a settleable floc.

Softening preceded coagulation during the dry season plant test. Although softening removed 20% of the DOC and 43% of the color, no additional removal of color and DOC was realized by the succeeding alum or PACl coagulation with softening than if coagulation was used alone. The addition of 1.2 meq/L of alkalinity to the softening process reduced the calcium hardness to only 100-mg/L as $CaCO_3$ although the alkalinity and total hardness were balanced. The softening pH was varied from 9.5 to 10.6 with 10.0 achieving the lowest hardness.

During the wet season the DOC was reduced an average of 57% to 8.6-mg/L by PACl and an average of 58% to 7.0-mg/L by alum. The dry season DOC removals averaged 48% with reductions 10.1-mg/L for alum and averaged 39% with reductions to 12.1-mg/L for PACl. This lower reduction during the dry season was due to a higher coagulation pH caused by a hard $CaCO_3$ scale in the reactor. The THM's at the plant tap during the wet season averaged 235 μg/L for PACl and 248 μg/L for alum. The dry season THM's at the plant tap averaged 204 μg/L for PACl and 175 μg/L for alum which represented a decrease of 13% to 30% in the dry season. However, average TOX for the alum or PACl treated water was 710 and 714 μg/L during the wet season. The TOX average for the alum or PACl treated water decreased to 421 and 380 μg/L during the dry season. Color removal was similar to DOC removal exhibiting removals of 91 to 94% for PACl and alum during the wet season and 78% to 89% for PACl and alum during the dry season.

The DOC:THM:TOX ratios during the wet season averaged 1:32:94 and 1:16:39 during the dry season. The THM:TOX ratio did not vary as much as the DOC:THM:TOX ratio since both THM:TOX are directly effected by Cl_2 dose. The THM:TOX ratio was 1:2.9 and 1:2.4 during the wet and dry seasons, respectively.

The operation cost increased from $1.07/1000 gal to $1.16/1000 gal when the maximum alum dose was used during the wet season. This was an 8% cost increase which decreased plant THM's by 72%. The operational cost during the wet season for PACl was $1.62/1000 and similar THM's reductions were obtained. The cost of producing water was less during the dry season due to relatively constant labor and power costs and increased production. The plant costs were increased from $0.60/1000 gal to $0.78/1000 gal using the maximum alum dose. This was a cost increase of 23% and reduced THM's by 39%. Dry season PACl costs were $1.02/1000 gal and decreased THM's by 29%.

It should be noted that no consumer complaints directed at the treatment process were received by Lee County during PACl/alum investigation at the Olga plant. Some complaints, as always, were received, but these were no more than normally would be expected for a utility serving 35,000 residents.

6.3 Sequential Treatment

6.3.1 Introduction

The purpose of this phase of the project was to evaluate the effectiveness of using lime softening and alum coagulation in series as a means to improve finished water quality. The AID site was chosen for this evaluation based on the raw water quality and the existence of two coagulation and clarification units on the site which were made available by the utility for the tests.

6.3.2 Sequence

The unit processes, raw water quality, and treatment processes utilized at AID have been described in detail elsewhere therefore only a brief description will be given here. Raw water is pumped from several shallow ground water wells and through an aeration tower into a raw water storage tank. Water from the storage tank is fed by gravity to the lime-softening unit. This lime-softening clarifier, as shown in Figure A-8, is rated at 1 MGD total flow and has a surface area of 694 ft^2.

The second clarification unit, as shown in Figure A-9, which in these tests were used for alum coagulation, has a rated capacity of 0.65 MGD for lime softening and a surface area of 451 ft^2. The lime-softening unit effluent was made to flow to the alum coagulation unit by connecting the two units with a 6" PVC line. A pump, as shown in Figure A-10, was installed between the two units because there was inadequate difference in evaluation between them. Figure 66 shows the flow of water through both plants and the sampling points when the plant was in operation. The dashed lines show the normal flow with the plants operating in parallel. The chemical feeds shown include sulfuric acid which was used to reduce pH of the pH 11.0 lime softened effluent. Alum and polymer were fed into the center of the alum coagulation reactor. Soda ash was used to increase the pH of the alum coagulated water prior to filtration. The chemical feeders are shown in Figure A-11.

During the first weeks of operation, the two plants were started and shut down every day of operation. As a result, it was difficult to stabilize the two processes. This problem was resolved during the study by running the lime-softening unit continuously. When the alum coagulation unit was shut down for the night, flow from lime-softening unit was diverted back to its normal path through the filters and into storage.

6.3.3 Jar Tests

Jar tests were performed using AID raw water. The purpose of these jar tests was to determine the optimum sequence of treatments, estimate chemical doses, pH and expected results of the plant scale tests.

In the first set of tests, raw water was treated using four different treatment sequences as shown in Figure 67. In the first sequence, lime softening at pH 10.3 followed by alum coagulation, hydrated lime was used to raise the pH of the raw water to pH 10.3. Following flocculation and settling, the supernatant was decanted and a sample held back for later analysis. The remainder of the supernatant was treated by coagulation with 90-mg/L of alum. Following flocculation and settling, the clarified supernatant was filtered through a glass fiber filter (Whatman GF/A) and saved for analysis. A second sequence was similar to the first, except that alum coagulation (90-mg/L alum) preceded lime softening at pH 10.3. In the third treatment sequence, lime softening was performed at pH 11.0 instead of 10.3. Following lime softening, alum coagulation was performed with pH adjustment to 5.5 using sulfuric acid. In the last treatment sequence alum coagulation at pH 5.5 preceded lime softening at pH 11.0.

A second set of jar tests was conducted, based on the results of the first jar tests. In these tests, lime softening preceded alum coagulation. Lime softening was performed at either pH 10.3 or 11.0 followed by alum coagulation at varying alum doses (29, 58, 117 and 175-mg/L as alum) and pH (5.0, 5.5, 6.0, 6.5 and 7.0). Note: Alum doses were originally selected as equivalent Al_2O_3. Expressed on this basis, the alum doses correspond to 5.0, 10.0, 20.0, and 20.0-mg/L. In order that the results relate more easily to actual operating conditions, these doses were converted to equivalent doses as $Al_2(SO_4)_3 \cdot 14\ H_2O$. This conversion is the reason for the doses used in the jar test and in the plant tests).

Results of the first set of jar tests are given in Table 59. For lime softening at pH 10.3 followed by alum coagulation, DOC reduction was 46%

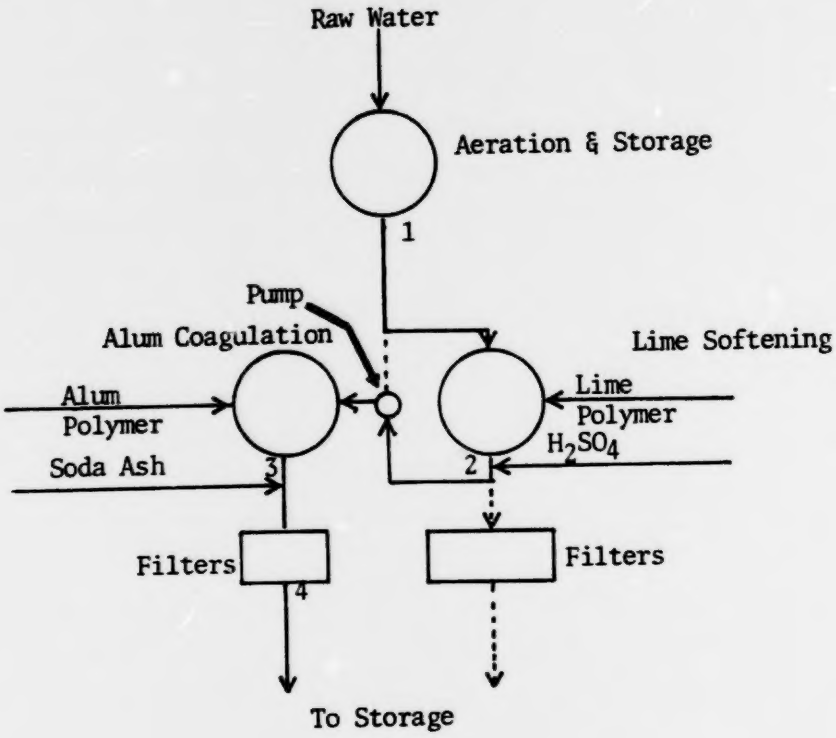

Raw Water
Aeration & Storage
1
Pump
Alum Coagulation
Lime Softening
Alum
Polymer
Soda Ash
3
Lime
Polymer
H_2SO_4
2
Filters
4
Filters
To Storage

Figure 66 Flow schematic for sequential treatment plant tests at the Acme Water Treatment Plant. Sampling points are 1 raw, 2 softened, 3 alum coagulated, and 4 filtered.

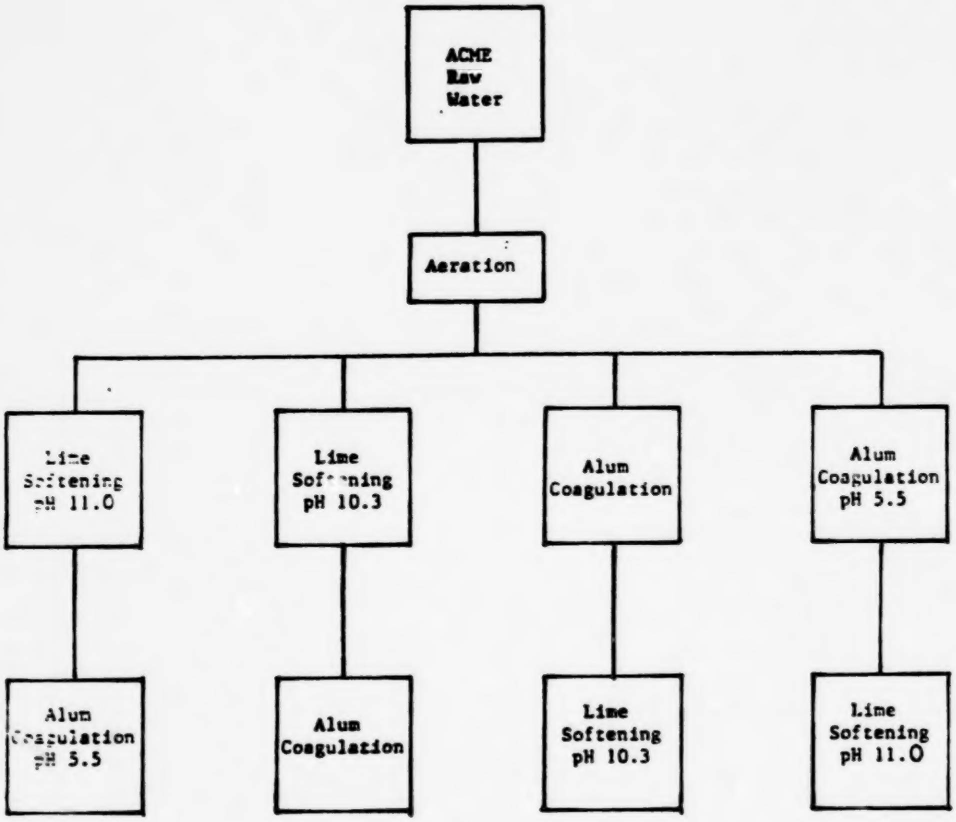

ACME
Raw
Water
Aeration
Lime
Softening
pH 11.0
Lime
Softening
pH 10.3
Alum
Coagulation
Alum
Coagulation
pH 5.5
Alum
Coagulation
pH 5.5
Alum
Coagulation
Lime
Softening
pH 10.3
Lime
Softening
pH 11.0

Figure 67 Sequential treatment jar test plan.

relative to the raw water, while the reverse sequence resulted in 43% DOC reduction. Thus the sequence of treatment had little impact on the overall DOC reduction. The DOC reduction by individual operations was affected, however, by the order in which they were performed. When alum coagulation with no pH adjustment preceded lime softening, only 16% of the DOC in the raw water was removed. However, when alum coagulation followed lime softening at pH 10.3, DOC was reduced by 24% relative to the softened water. Lime softening alone at pH 10.3 reduced DOC 29% relative to the raw water. Following alum coagulation, lime softening reduced DOC by 32 % relative to the coagulated water. Despite these differences, the two sequences resulted in nearly the same final DOC, 8.22-mg/L for lime softening followed by alum and 8.65-mg/L for alum coagulation followed by lime softening.

Using pH 11.0 for softening and controlling the alum coagulation pH at 5.5 resulted in 62% DOC reduction vs. 59% for the opposite sequence. These higher reductions were basically the result of optimizing both the lime-softening and the alum coagulation processes with respect to pH. Increasing the pH of lime softening from 10.3 to 11.0 resulted in a 46% reduction in DOC relative to the raw water. This reduction was approximately 23% greater than that achieved by lime softening at pH 10.3. Alum coagulation at pH 5.5 alone resulted in a 44% DOC reduction which was 28 percent better than coagulation with no pH adjustment. Combining alum coagulation at pH 5.5 and lime softening at pH 11.0 resulted in overall DOC reductions of approximately 60% regardless of sequence. This represented about 28% improvement in DOC removal relative to the non-optimized sequence. In both cases, the sequence of operations did not appear to significantly effect the process performance for the reduction.

THM formation potential was 342 μg/L and 351 μg/L for lime softening at pH 10.3 and 11.0 respectively. The THM's for alum coagulation without controlling pH and alum coagulation at pH 5.5 were 424 μg/L and 260 μg/L respectively (39% improvement). The THM's for pH 10.3 softened, alum coagulated water was 239 μg/L, while the THM's for the opposite sequence was 301 μg/L. For water softened at pH 11.0 followed by pH 5.5 alum coagulation, THM's was 217 μg/L, while the THM's for the opposite sequence was 167 μg/L. Thus optimizing pH resulted in a 9% THM's decrease for lime softening followed by alum coagulation and a 45% decrease in THM's for alum coagulation followed by lime softening. The difference in percent improvement appears to be due to the improvement in the effectiveness of alum coagulation without pH adjustment following lime softening. Lime softening reduces alkalinity and therefore the alum applied can more effectively lower pH during coagulation in this instance.

TOX formation potential was 1283 μg/L in the pH 10.3 softened water and 960 μg/L in the pH 11.0 softened water. This represents a 25% decrease in TOXFP by increasing softening pH to 11.0 from 10.3. TOXFP for alum coagulation without pH control was 1212 μg/L and was decreased to 884 μg/L (31%) by controlling the coagulation pH to 5.5. For pH 10.3 softened water, TOXFP was reduced to 832 μg/L when alum coagulation followed lime softening. When alum coagulation preceded softening, sequential treatment reduced TOXFP to 954 μg/L. When lime softening at pH 11.0 is followed by

Table 59. Results for Sequential Treatment Jar Tests Using Acme Raw Water

Treatment	DOC (mg/L)	TOXFP (μg/L)	THM's (μg/L)	Color (cpu)	Al (mg/L)	pH	TH (mg/L as $CaCO_3$)	CH (mg/L as $CaCO_3$)	Alk (mg/L as $CaCO_3$)
Acme Raw	15.21	1546	532	47	0.21	7.4	330	320	305
Aerated	14.76	1444	508	34	0.21	7.9	296	284	269
L.S. pH 10.3	10.75	1283	342	21	0.09	8.8	82	59	54
L.S. pH 10.3/Alum	8.22	832	239	8	0.10	6.6	140	90	45
Alum	12.73	1212	424	18	0.25	7.7	328	320	269
Alum/L.S. pH 10.3	8.65	954	301	5	0.18	10.5	136	133	80
L.S. pH 11.0	8.28	960	351	7	0.16	10.5	108	104	96
L.S. pH 11.0/Alum pH 5.5	5.83	603	217	4	0.14	6.8	178	166	17
Alum pH 5.5	8.56	884	260	8.0	0.30	6.0	336	320	38
Alum pH 5.5/L.S.pH 11.0	6.23	544	167	5.0	0.67	11.0	492	434	154

Alum dose = 15 mg/L Al_2O_3
L.S. = lime softened

alum coagulation at pH 5.5, TOXFP was reduced to 603 μg/L, while the opposite sequence reduced TOXFP to 544 μg/L. Thus optimizing pH of both processes resulted in 28% improvement for lime softening followed by alum coagulation and 43% for alum coagulation followed by lime softening. Despite the greater percent improvement for optimizing the alum/lime sequence, the difference between the two sequences for each set of pH conditions was less important (15% and 10%).

Color reduction by sequential treatments varied from 83% to 92% of the raw water color. A 69% improvement in color reduction was achieved by increasing lime softening pH from 10.3 to 11.0 (21 cpu vs 6.5 cpu) and reducing the alum coagulation pH to 5.5 resulted in a 56% improvement in color removal (18 vs 8 cpu). Optimizing pH for both processes resulted in a 38 percent improvement in color reduction for the lime-softening/alum sequence, however no improvement was achieved by optimizing pH in the alum coagulation/lime-softening sequence.

Aluminum residuals were low for all treatments except for alum coagulation at pH 5.5 followed by lime softening at pH 11.0 in which some of the precipitated aluminum floc was redissolved by the high lime-softening pH. This resulted in an aluminum residual of 0.67-mg/L which was more than three times the background level. Alum coagulation alone, both with and without controlling pH at 5.5, also increased aluminum slightly above background levels.

Total hardness was reduced to acceptable levels for all treatments involving lime softening at pH 10.3. When lime softening at this pH was used together with alum coagulation, the final hardness was higher than when softening alone was used (140 and 136-mg/L $CaCO_3$ vs 82-mg/L $CaCO_3$). For pH 10.3 lime softening followed by alum coagulation, this is probably due to redissolving $CaCO_3$ floc. For alum coagulation followed by pH 10.3 softening, however, it appears that the lime dose was greater than necessary to achieve a final pH of 10.3. This may have been due to alkalinity lost as CO_2 during alum coagulation. For lime softening at pH 11.0 followed by alum coagulation at pH 5.5, the total hardness was 173 μg/L $CaCO_3$. However, for alum coagulation at pH 5.5 followed by lime softening at pH 11.0 the final hardness was 492-mg/L $CaCO_3$. Higher hardness residuals were expected for softening at pH 11.0 when neither CO_2 nor Na_2CO_3 was used to precipitate the excess lime. However, for this last sequence the residual hardness and alkalinity indicates that the lime dose was too high.

From the first set of jar tests, it was concluded that lime softening followed by alum coagulation would be an easier sequence to control and would require less additional chemical applications in the form of acid, soda ash, CO_2, etc. It was suggested that additional benefit could be gained from high pH softening so the following jar tests and plant tests were designed to determine if this was the case.

Analytical results from the second set of sequential treatment jar tests are given in Tables 60 and 61. DOC in the raw water was 14.6-mg/L.

TABLE 60. AID Sequential Treatment Jar Test Data

Softening pH	Alum Dose* (mg/L)	Coagulation pH	DOC (mg/L)	TOXFP (ug/L Cl)	THMFP (ug/L)	Color (cpu)
Raw	0	--	14.61	1356	505	38.0
10.3	0	--	10.12	552	185	23.0
10.3	29	5.0	7.23	388	190	9.0
		5.5	6.74	469	200	6.5
		6.0	6.78	494	130	6.5
		6.5	7.06	510	125	7.0
		7.0	7.24	447	160	8.0
10.3	58	5.0	5.75	350	240	5.0
		5.5	5.19	418	250	3.0
		6.0	6.15	433	135	5.0
		6.5	6.08	422	105	4.5
		7.0	6.17	482	125	5.0
10.3	117	5.0	4.87	391	205	3.5
		5.5	4.78	337	100	3.0
		6.0	4.96	341	98	3.0
		6.5	5.44	401	100	3.5
		7.0	6.14	460	150	5.5
10.3	175	5.0	4.40	411	240	5.5
		5.5	4.67	287	135	3.5
		6.0	5.13	337	140	4.0
		6.5	5.34	361	125	4.0
		7.0	7.06	483	160	4.0

*mg/L $Al_2(SO_4)_3 .14H_2O$

TABLE 61. AID Sequential Treatment Jar Test Data

Softening pH	Alum Dose* (mg/L)	Coagulation pH	DOC (mg/L)	TOXFP (ug/L Cl)	THMFP (ug/L)	Color (cpu)	Al (mg/L)
Raw	0	0	14.6	1356	505	38	0.1
11.0	0	0	8.70	475	165	14	0.33
11.0	29	5.0	6.76	521	110	7.5	0.08
		5.5	6.42	424	105	6.5	0.08
		6.0	6.56	532	90	6.0	0.12
		6.5	6.30	473	160	6.0	0.06
		7.0	8.04	531	155	9.0	0.12
11.0	58	5.0	5.27	345	95	4.0	0.08
		5.5	5.80	406	60	4.0	0.13
		6.0	6.06	421	140	5.0	0.10
		6.5	5.81	430	160	5.5	0.13
		7.0	5.80	444	150	5.0	0.13
11.0	117	5.0	4.97	345	90	3.0	0.06
		5.5	5.64	401	80	4.0	0.07
		6.0	4.33	293	100	3.0	0.06
		6.5	6.18	430	110	5.0	0.10
		7.0	5.48	450	130	5.0	0.06
11.0	175	5.0	4.96	287	90	3.5	0.06
		5.5	4.59	277	75	2.0	0.06
		6.0	4.77	331	75	3.5	0.06
		6.5	6.08	413	140	5.0	0.10
		7.0	5.02	408	140	3.5	0.07

*mg/L $Al_2(SO_4)_3 \cdot 14H_2O$

This value was slightly lower that the raw DOC from the first set of jar tests and from the plant scale tests and was a result of using different raw water wells on different days. Lime softening alone at pH 10.3 resulted in a DOC of 10.1-mg/L. This was a 31% reduction in DOC relative to the raw water. Lime softening at pH 11.0 reduced DOC to 8.70-mg/L which represents a 14% DOC reduction relative to the pH 10.3 softened water DOC and 40% reduction relative to the raw water.

Using alum coagulation in series with lime softening improved overall DOC removal. For alum doses lower than 58-mg/L in the water softened at pH 10.3, DOC reduction was insensitive to coagulation pH (see Figures 68 and 69). DOC removal for coagulation using 29-mg/L alum was approximately 30% for pH 10.3 softened water and slightly less using pH 11.0 softened water, regardless of the coagulation pH. For alum doses greater than 59-mg/L, the percent DOC reduction becomes more dependent on coagulation pH. Maximum DOC reduction by alum coagulation occurs when coagulation pH is maintained less than 6.0. Figures 68 and 69 show overall DOC reduction, relative to the raw water, is maximized at about 70% for alum doses of 175-mg/L and pH 5.5 for either softening pH, however the DOC of samples coagulated at pH 5 to 6 following softening at pH 11.0 were approximately 0.5-mg/L (3%) lower than similar samples softened at pH 10.3. Most of the additional DOC removed by softening at pH 11.0 would be removed by alum coagulation following softening at pH 10.3. The increased DOC removal by softening at pH 11 vs. 10.3 and following coagulation is small.

The THM's were always measured at 25°C, 48 hr. 15-mg/L Cl_2 and pH 7.5. THM's for the pH 10.3 and pH 11.0 softened waters were 185 and 165 μg/L. The 11 percent reduction in THM formation at the high pH is similar to the 14 percent decrease in DOC and TOX formation observed, but considerably less than the 40 percent decrease in color that resulted from increasing the lime-softening pH.

Further reductions in THM formation were achieved by alum coagulation after lime softening. Figures 70 and 71 show that the most significant improvements were achieved when the alum coagulation pH was held less than 6.5, and that practically no advantage over softening was obtained when coagulation pH equaled 7.0. Although there was considerable variability in the data, it appears that alum coagulation between pH 5.5 and 6.0 can reduce THM formation by as much as 50 percent or more for either softening pH. One major difference was at pH 5.0, where alum coagulation reduced THM formation in the water softened a pH 11.0 almost twice as much as water softened at pH 10.3.

The average difference in THM formation for water softened at 10.3 and 11.0, followed by alum coagulation was 43 μg/L which was significant. Thus lime softening at pH 10.3 followed by alum coagulation resulted in significantly higher THM's than lime softening at pH 11.0 followed by alum coagulation. In four samples, THM's were lower for pH 10.3 softened water than for pH 11.0 softened water. However, these differences were small (10 μg/L or less) compared to the average difference for all samples.

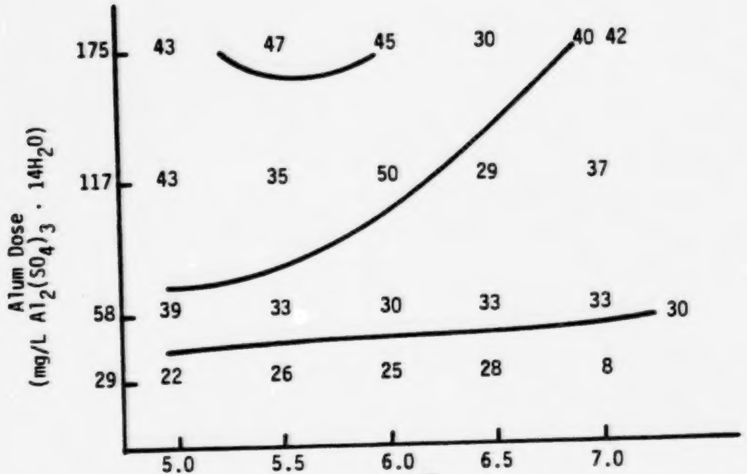


Figure 68 Isopleth for percent DOC reduction by Alum coagulation following lime softening at pH 10.3. Raw DOC = 14.6 mg/L, Softened DOC = 10.1 mg/L.

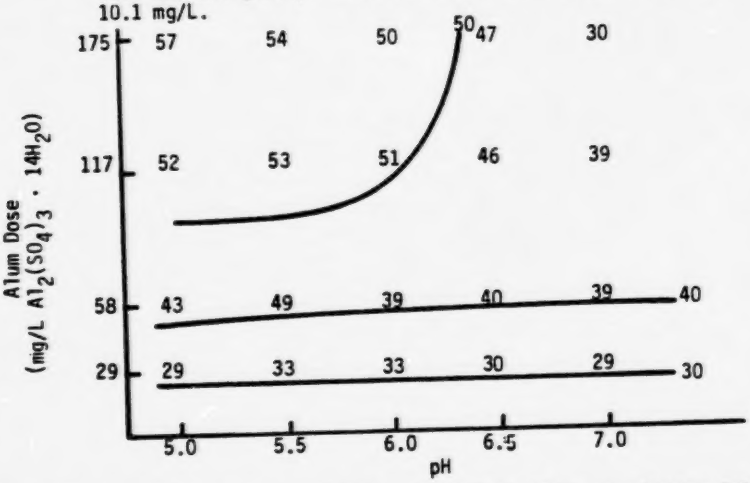


Figure 69 Isopleth for percent DOC reduction by Alum coagulation following lime softening at pH 11.0. Raw DOC = 14.6 mg/L, Softened DOC = 8.7 mg/L.

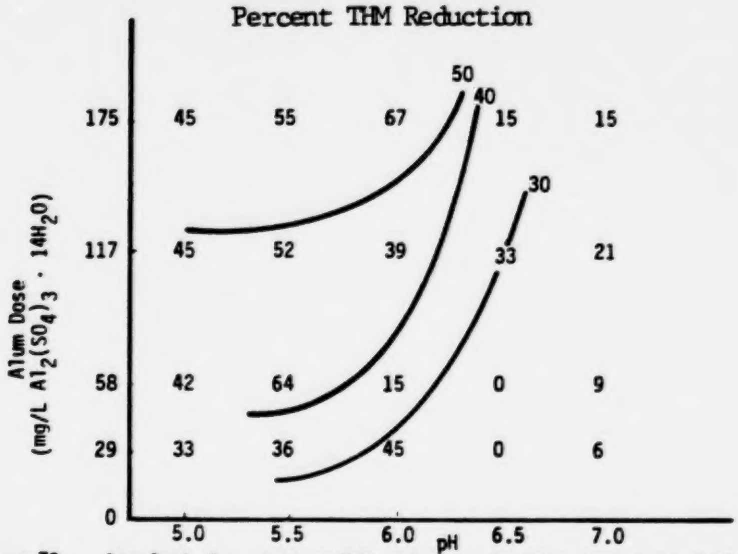


Figure 70 Isopleth for percent THM reduction by Alum Coagulant following lime softening at pH 11.0. Raw THMFP = 505 ug/L. Softened THMFP = 165 ug/L.

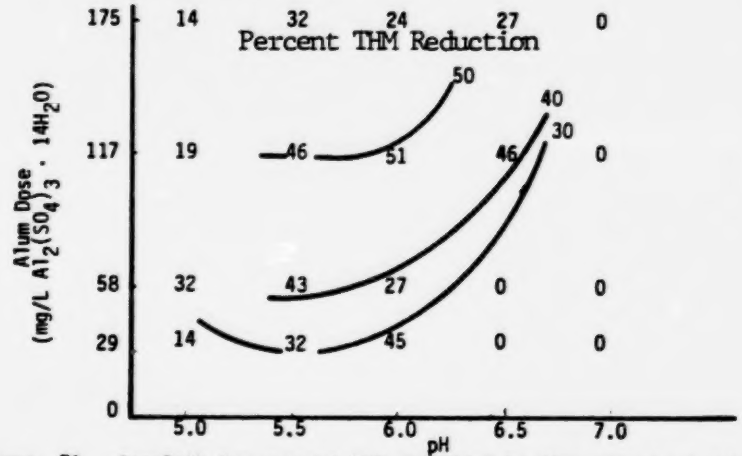


Figure 71 Isopleth for percent THM reduction by Alum coagulant following lime softening at pH 10.3. Raw THMFP = 505 ug/L. Softened THMFP = 185 ug/L.

TOX values for the AID sequential treatment jar tests varied from 277 to 552 μg/L. Lime softening resulted in TOX concentrations of 552 and 475 μg/L for pH 10.3 and 11.0 respectively. Thus a 14% decrease in TOX formation can be attributed to increasing the softening pH from 10.3 to 11.0. This contrasts with a 14% decrease in DOC, an 11% decrease in THM formation and a 40 percent decrease in color.

Further reductions in the TOXFP by alum coagulation are shown in Figure 72. These constant percent removal graphs show that over 40 percent additional TOX reduction was achieved using alum coagulation at pH 5.5 and doses as low as 117-mg/L. For all alum doses, the reduction in TOX formation decreased significantly as the coagulation pH increased from 6.0 to 6.5. The percent reduction in TOX formation for softened water treated with alum at doses lower than 58-mg/L was not consistently affected by coagulation pH although slightly higher percent reductions were noted for coagulation at pH 5.0 at these low alum doses.

Comparing the differences in TOXFP between water softened at pH 10.3 and 11.0 for each combination of alum dose and coagulation pH, resulted in an average difference of 2.5 μg/L. Thus no significant difference in the overall TOX formation between the two softening pHs followed by alum coagulation is evident. For the twenty corresponding values of alum dose and coagulation pH, there were six instances where the TOXFP for pH 10.3 softened water was less than pH 11.0 softened water and 14 instances where the TOXFP for pH 11.0 softened water was less than pH 10.3 softened water. This indicates that at higher alum doses the water softened at pH 11.0 had lower TOX values than that softened at pH 10.3. These results also show that better performance results from alum coagulation at low alum doses following softening at pH 10.3 compared to pH 11.0. This result suggests that alum coagulation requires some buffering capacity in the form of carbonate alkalinity in order to be effective for TOX precursor removal.

Color in the raw water for the AID jar tests was 38 cpu. Lime softening at pH 10.3 reduced color to 23 (40% reduction)-mg/L and at pH 11.0 reduced color to 14 (63% reduction) cpu. This 40 percent increase in color reduction at the high pH was greater than the improvement that resulted in other parameters, DOC, THMFP, and TOXFP, from raising the softening pH.

Further significant color removal was obtained by alum coagulation following softening at pH 10.3 or 11.0. For water that had been softened at pH 10.3, color following alum coagulation varied from 3.0 to 9.0 cpu (87 to 61 percent further reduction). This represented an overall color reduction of over 92 percent. For water that had been softened at pH 11.0, alum coagulation reduced color from 14 to between 2 and 9 cpu (36 to 86% reduction). Overall color reduction following alum coagulation for water softened at pH 11.0 varied from 76 to 95%.

Figures 73 and 74 show percent color reduction by alum coagulation following lime softening at pH 10.3 or pH 11.0, in both cases, the majority of color reduction is achieved at low alum doses. For water softened at pH 10.3 between 60 and 70 percent color reduction is achieved with an alum

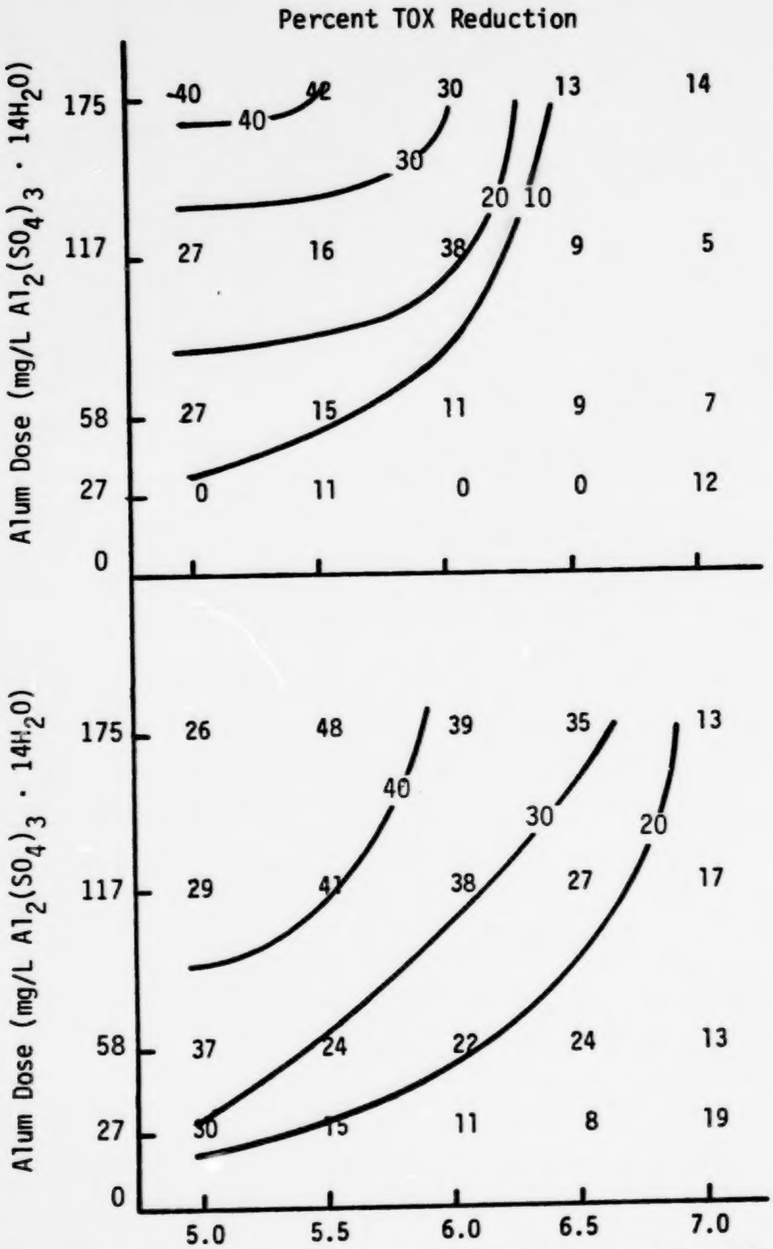


Figure 72 TOX Isopleths by Alum coagulation following lime softening at pH 11.0 (Top) and 10.3 (bottom). Raw water TOX = 1356, softened water. TOX = 475 (pH 11.0)
TOX = 552 (pH 10.3)

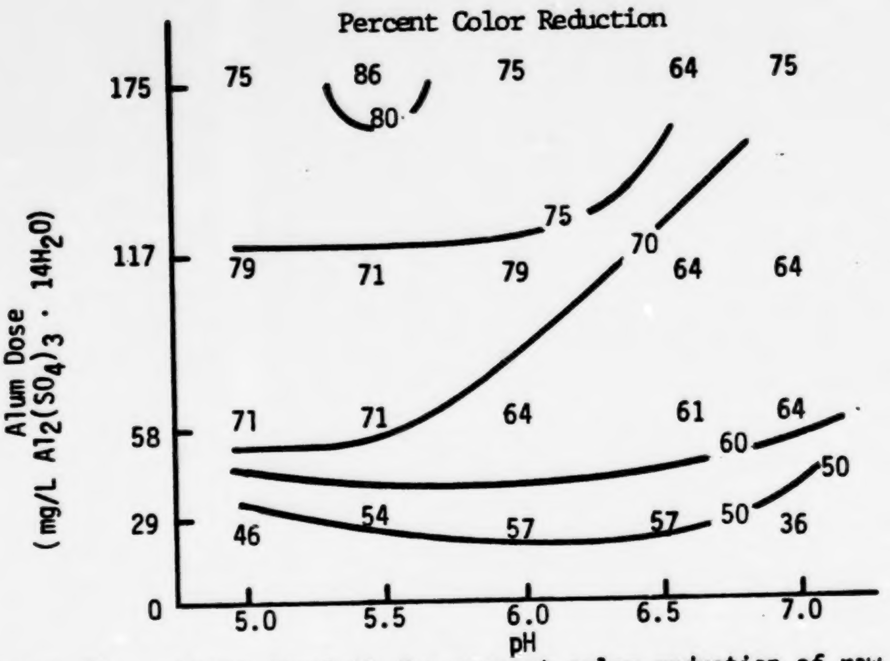


Figure 73 Color Isopleth for percent color reduction of raw water color treated by Alum coagulation following lime softening at pH 11.0. Raw water color of 38 cpu, softened color of 14.

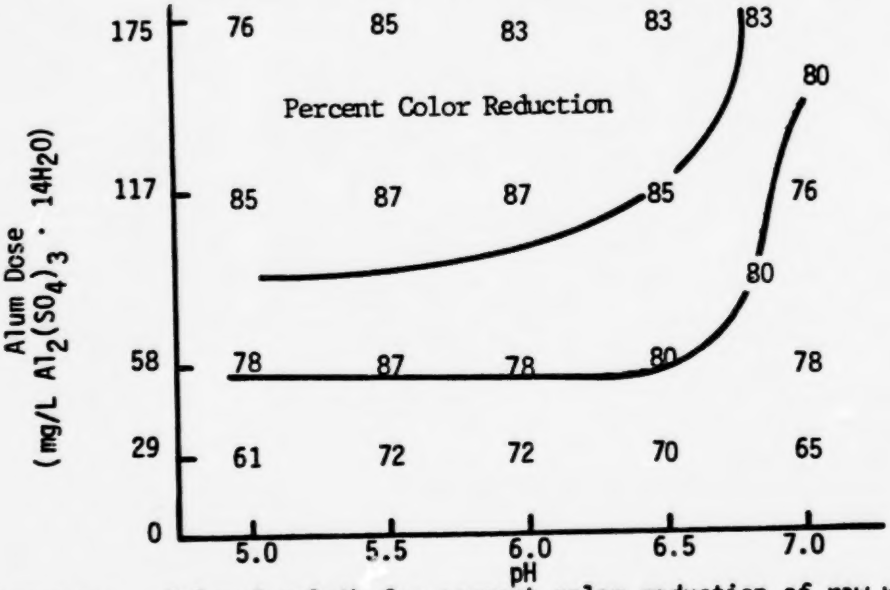


Figure 74 Color Isopleth for percent color reduction of raw water color treated by Alum coagulation following lime softening at pH 10.3. Raw water color of 38 cpu, softened color of 23.

dose of 29-mg/L and for water softened at pH 11.0 40-50 percent color reduction is achieved with this same alum dose. Increasing alum doses in both cases resulted in significantly diminished additional color reductions. Maximum color reduction appears to occur for an alum dose of 117-mg/L.

Color reduction by alum coagulation following lime softening appeared to be more effective at pH values less than 6.5, particularly when lime softening at pH 11.0 preceded coagulation. The average difference in color between water softened at pH 10.3 and 11.0 was 0.15 cpu for identical alum doses and coagulant pH. Thus no significant difference in overall color removal was observed between the two different pHs for lime softening.

6.3.4 Plant Tests

Plant scale testing at the AID plant ran from March 29 through May 5, 1985. During this time, the sequence of treatments used was lime softening followed by alum coagulation. Sampling points during the study include the raw water, softened effluent, alum coagulated effluent, and filtered effluent. Chlorination of the samples was accomplished in the laboratory using filtered samples. This was necessary, rather than sampling after chlorination in the plant, because no sampling point was available that would isolate treated water from stored water.

Flow through the plant was controlled at 175 GPM. This flow provided a surface loading rate of 0.4 gpm/ft^2 in the alum coagulation accelerator and 0.25 gpm/ft^2 in the softener. Surface loading on the filters was 1.5 gpm/ft^2. The loading rates were controlled at these levels in order to prevent solids carry over from the alum coagulation accelator. The arrangement of piping in the plant precluded using the larger accelator for alum coagulation, which would have been desirable.

Lime was fed from a conventional lime slaker to the lime-softening accelator. Because the flow was low, lime feed rates were difficult to control. Typically, the lime feed belt operated at approximately 4% of it's maximum speed. Even small perturbations in plant flow rate or lime feed were noticeable in the lime-softening process.

Acid was fed, as necessary, in the effluent of the lime softener. This was necessary only during periods of high pH lime softening. Acid feed rates were seldom over 10-20 mL/min. Alum was mixed using 2 lb of dry powdered alum per gallon of water. This mixture was fed by positive displacement pump to the center of the alum coagulation accelator. An organic polymer (Olin 5011) was fed along with alum at a dose of 0.4-mg/L. After the alum coagulation accelator, the settled water pH was adjusted to 8.0 with sodium carbonate and filtered.

Plant scale operation was intermittent, usually between 8 and 16 hours per day. In order to avoid the problems of frequent start ups, as previously mentioned, lime softening continued on a 24 hour basis, with the softened water being chlorinated, filtered and sent to storage when the alum coagulation unit was not operating.

Data for the plant scale sequential operation are given in Tables A-12 and A-13. Table A-12 gives water quality parameters for the effluent of the lime-softening reactor. The percent reduction for DOC and color are relative to the raw water. In Table A-13, the same parameters are described for the alum coagulation reactor effluent and percent removals are related to both the raw and softened water. In addition to water quality parameters, lime and alum doses are given for each of the operating conditions.

6.3.4.1 DOC

DOC concentrations for each point in the sequential treatment operation are shown in Figure 75. DOC in the raw water varied from 21.4-mg/L to 14.7-mg/L. This variability resulted from the operation of different wells at different times. After an initial DOC drop in the raw water was noted an effort was made to supply the raw water reservoir using the same wells in order to stabilize raw water quality. DOC reduction in the softening process varied from 30 to 54 percent of the raw water DOC, during operation at pH 10.3. DOC reductions of over 50% on 3-29 and 3-30 were artifacts of the high raw water DOC and when these values are removed, DOC reduction varied from 30 to 41 percent during softening at pH 10.3 and the average DOC reduction was 38%. During the jar tests, DOC reduction by lime softening was 30% and the resulting DOC was approximately 10-mg/L.

At pH 11.0 some improvement in DOC removal was noted. At this pH, DOC was reduced by 43-52% in softening, resulting in values between 7.66 and 9.45-mg/L. Average percent removal was 48% at the higher pH and the average DOC value was 8.5-mg/L. Thus, increasing the softening pH from 10.3 to 11.0 resulted in roughly 10% increase in DOC removal.

Further DOC reduction was accomplished by coagulation. The average additional DOC reduction afforded by alum coagulation after lime softening at pH 10.3 varied from 21 to 50%, based on the softened water DOC. The alum coagulation pH during these tests was controlled by the alum dose which varied from 68 to 109-mg/L. Optimum performance was achieved with the highest alum dose and lowest pH, 109-mg/L and 5.0 respectively. Under these conditions, DOC was reduced an additional 50% based on the softened water DOC for an overall DOC reduction of 70% based on the raw water DOC. These values correspond well with DOC removal in the sequential treatment jar tests. For water softened at pH 10.3 and coagulated at pH 5.0 with an alum dose of 117-mg/L, overall DOC reduction was approximately 67%. For other values of pH and alum dose, the jar tests and plant test results were generally comparable.

For softening at pH 11.0 followed by alum coagulation, performance in the plant tests was slightly better than that in jar tests. Overall DOC reduction in the jar tests was approximately 60% but in the plant tests, DOC was reduced by an average of 70%. As mentioned above, the jar test data did not indicate any difference in overall DOC removal efficiency between softening at pH 10.3 or 11.0 followed by alum coagulation. This was also found to be the case in plant tests. It should be noted that in

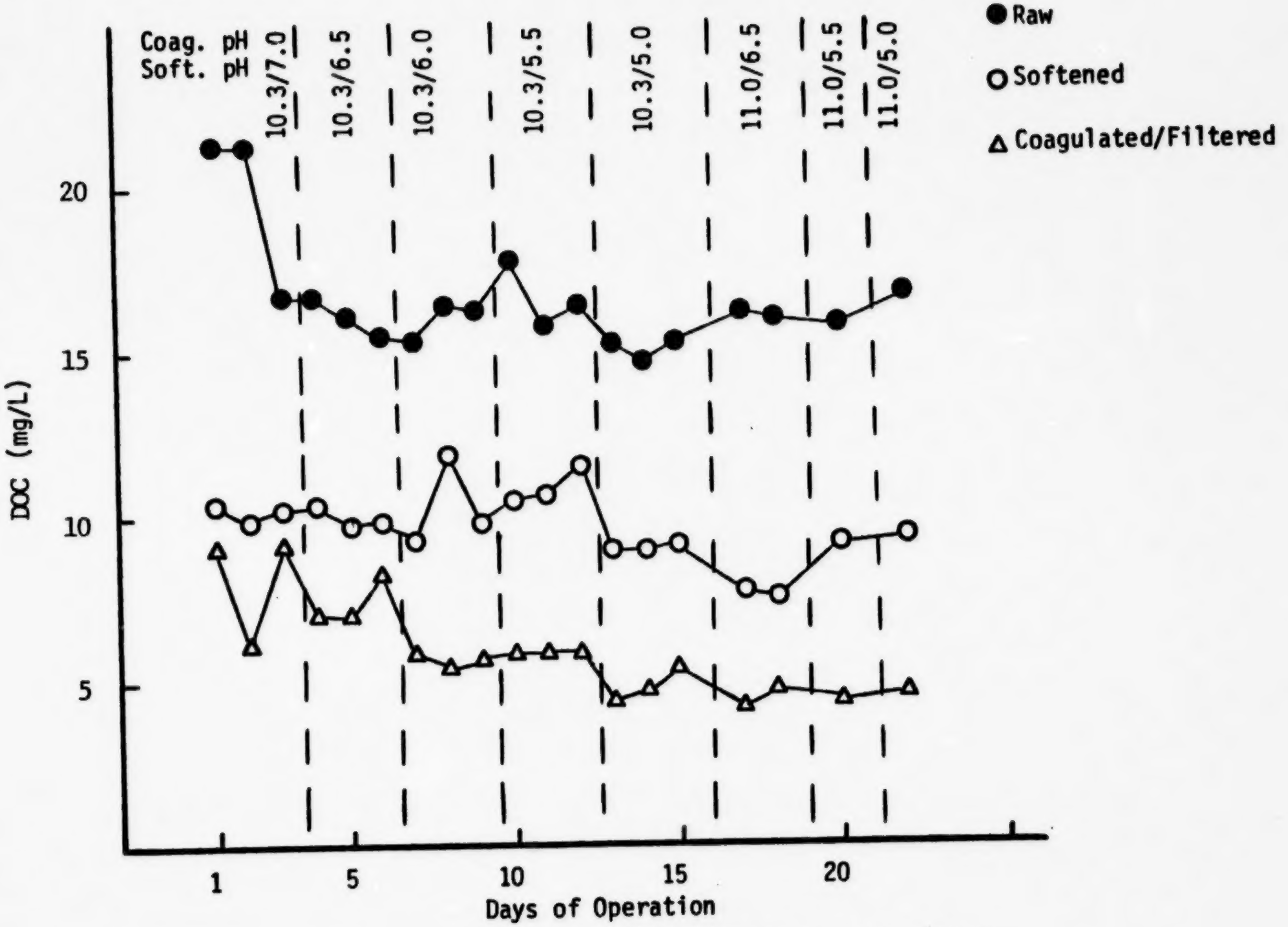


Figure 75 Raw, softened and coagulated DOC during Sequential Treatment Plant Test.

both cases, the control of the pH 11.0 processes was more difficult due to the loss of carbonate alkalinity in the high pH process.

Table 62 shows the percent DOC reduction achieved in the plant tests compared to the results achieved under similar conditions in the jar tests. This table shows that in general, agreement was good between the jar tests and the plant tests. Agreement was best for the softening at pH 10.3 and was worse for softening at pH 11.0. This lack of agreement for high pH softening results from difficulty in controlling the two processes when carbonate alkalinity is lost.

6.3.4.2 THM

Trihalomethane formation potential was measured in samples taken from the lime-softening effluent and the filtered alum coagulation effluent. These samples were chlorinated as described elsewhere in this report.

THM's varied over the course of the study as described in Figure 76. THM's in the softened water varied between 215 μg/L and 472 μg/L with an average of 339 μg/L. These values are similar to the THM values found in the normal operation of the plant for lime softening. When lime softening at pH 11.0 was performed, THM's increased to an average of 445 μg/L. This result was unexpected since softening at high pH had decreased THM's in other studies both at AID and other locations. In the jar tests, increasing the softening pH from 10.3 to 11.0 resulted in a decrease in THM formation from 185 to 165 μg/L or approximately 11%. This result corresponded to the 10% reduction in DOC also observed in the jar tests.

When alum coagulation was used in series with lime softening, THM's were reduced by an additional 20 to 30% compared to the softened water THM's. THM's values for the finished water varied from 145 to 414 μg/L. The lowest THM values resulted from softening at pH 10.3 followed by coagulation at pH 6.5 with an alum dose of 76-mg/L. This result was different from what would be expected, based on the jar test results. Based on the jar test results, a further THM's reduction of 40 to 50 percent would be expected when alum coagulation at these conditions is used following lime softening. The higher alum doses and lower pH values did not result in lower THM formation potentials nor in higher percent reduction. This result was also contradictory to the jar test results. The most probable reason for these is that floc carry over in the alum coagulation accelator was often a problem. Frequent starting and stopping of the operation made the system difficult to stabilize even when the lime-softening unit was allowed to run for 24 hours per day. It is probable that this was the major cause of the problem, since the greatest discrepancy in jar test and plant test results occurred at the higher alum doses. This would suggest that the problem was basically one of improving clarification.

THM reductions were better for the high pH softening followed by alum coagulation. The greatest percent reduction was achieved by following lime softening with alum coagulation at pH 5.5 which reduced THM's by 45%

TABLE 62. Comparison of % DOC Reduction in Sequential Treatment Plant Tests vs Corresponding % DOC Reduction in Jar Tests

Softening pH	Coagulation pH	Alum Dose (mg/L)*	% DOC Reduction	
			Jar Test	Plant Test
10.3	7.0	68	58	58
	6.5	76	58	53
	6.0	83	62	64
	5.5	94	67	65
	5.0	109	67	70
11.0	6.5	109	60	71
	5.5	109	61	71

*mg/L $Al_2(SO_4)_3 \cdot 14H_2O$

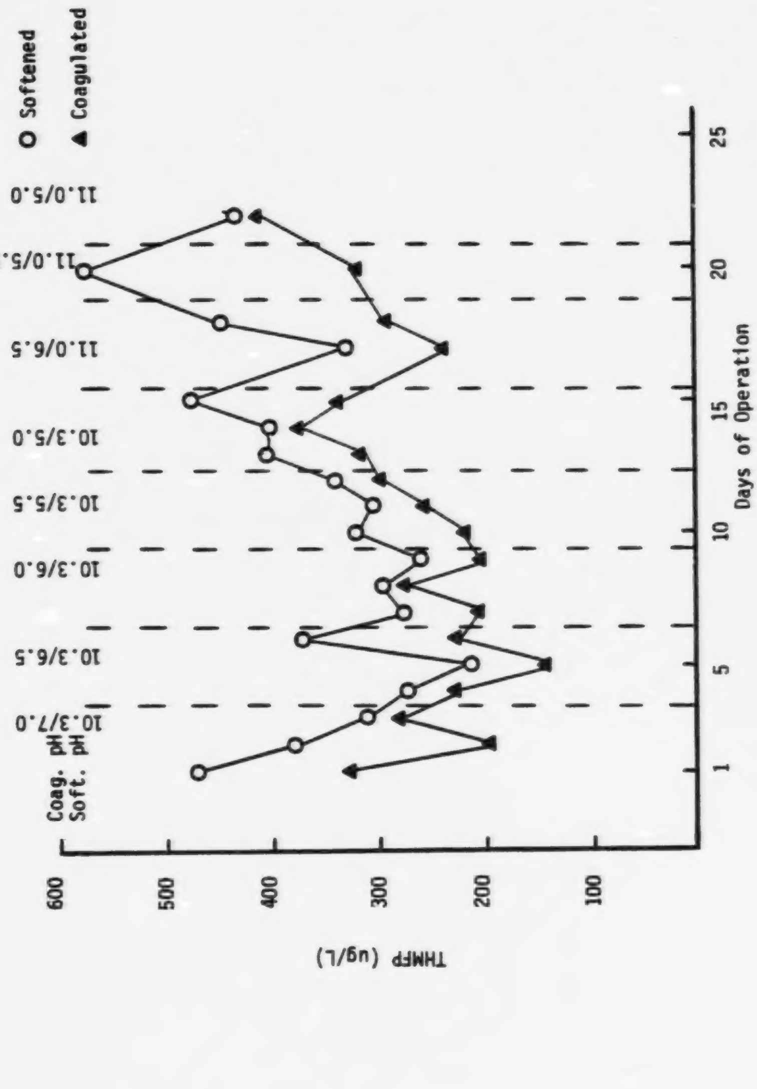


Figure 76 Softened and coagulated THMFP during Sequential Treatment Plant Test.

relative to the softened water. On the average, THM's were reduced by an additional 30% relative to the softened water.

THM reductions were not identical to the DOC reduction at the higher alum doses. DOC reduction of 46% was achieved at an alum dose of 94-mg/L while THM reduction was increased at the same dose by only 20 percent.

6.3.4.3 TOX

TOX analyses were conducted on samples from the same points as THM analyses. TOX in the pH 10.3 lime softened water averaged 796 µg/L. For the pH 11.0 softened water the average TOX was 793 µg/L, thus there was no significant difference in TOX resulting from softening at the two pH values (see Figure 77). These values were higher than obtained in the jar tests (552 µg/L and 475 µg/L). The reason for the difference could be attributed to raw water being taken from different wells. Although a pattern of generally higher values is followed for TOX and THM in the plant tests than in the jar tests, the treatment trends are generally similar for equivalent operating conditions.

TOX in the alum coagulated samples was reduced between 2 and 35% relative to the lime softened water. Significant differences were found between lime softened water TOX and alum coagulated water TOX for alum coagulation at pH 6.0 and 5.5.

TOX in the lime softened water varied over time with the TOX on the dates 4-16 through 4-21 being significantly less than the rest of the softened water samples. These samples also had slightly lower DOC values than the rest of the lime softened samples.

The lowest TOX values were obtained by alum coagulation at pH 6.0 following lime softening. Overall the average TOX in the pH 11.0 softened and alum coagulated water was greater than or equal to the pH 10.3 softened and coagulated water.

6.3.4.4 Color

Color for the raw, softened, and coagulated water is shown in Figure 78. Color in the raw water varied between 40 and 50 cpu for most of the test period. Lime softening at pH 10.3 reduced color to approximately 20 cpu (50-60%). Increasing the softening pH to 11.0 reduced color further to approximately 12 cpu. This color removal was increased by approximately 20% by increasing the softening pH to 11.0.

Alum coagulation further reduced color by approximately 50% to 80% relative to the softened water. In all cases, color was reduced to less than 15 cpu, but lowest color values were achieved by alum coagulation at 5.0 following lime softening at pH 10.3. For alum coagulation at higher values of pH, color was lower when pH 11.0 softening preceded coagulation. These data show that when alum coagulation is optimized, no advantage is gained for color removal by higher pH softening.

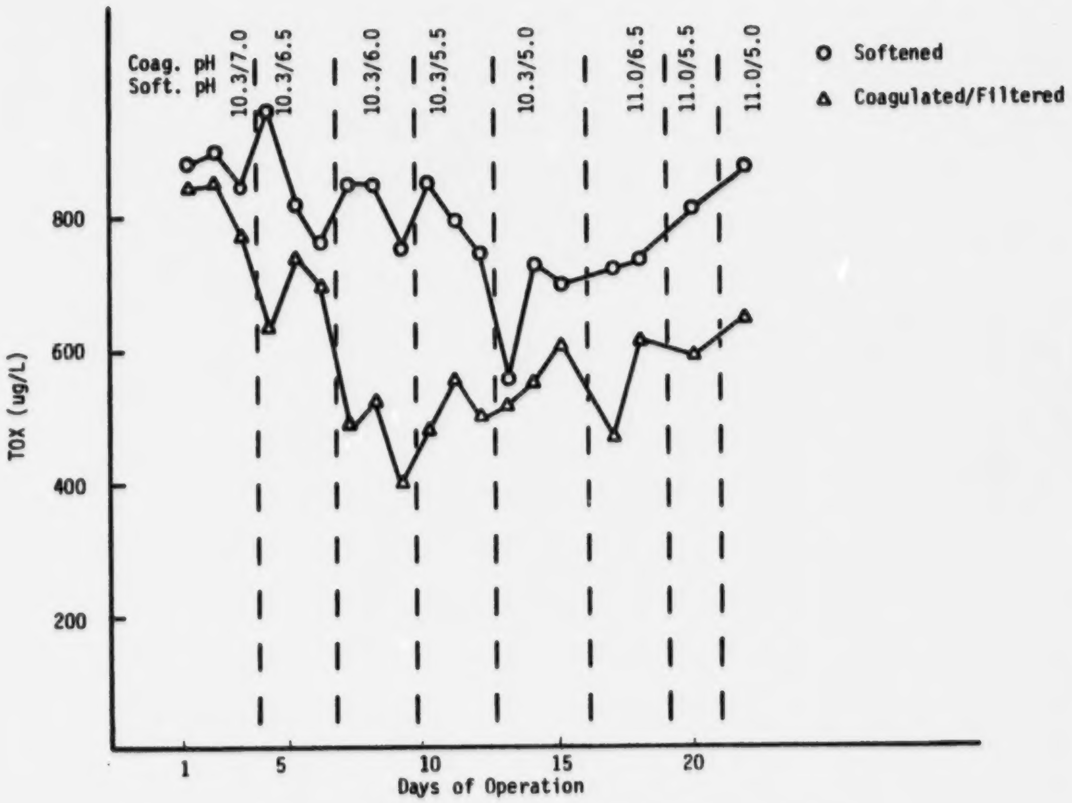


Figure 77 Softened and coagulated TOX during Sequential Treatment Plant Test.

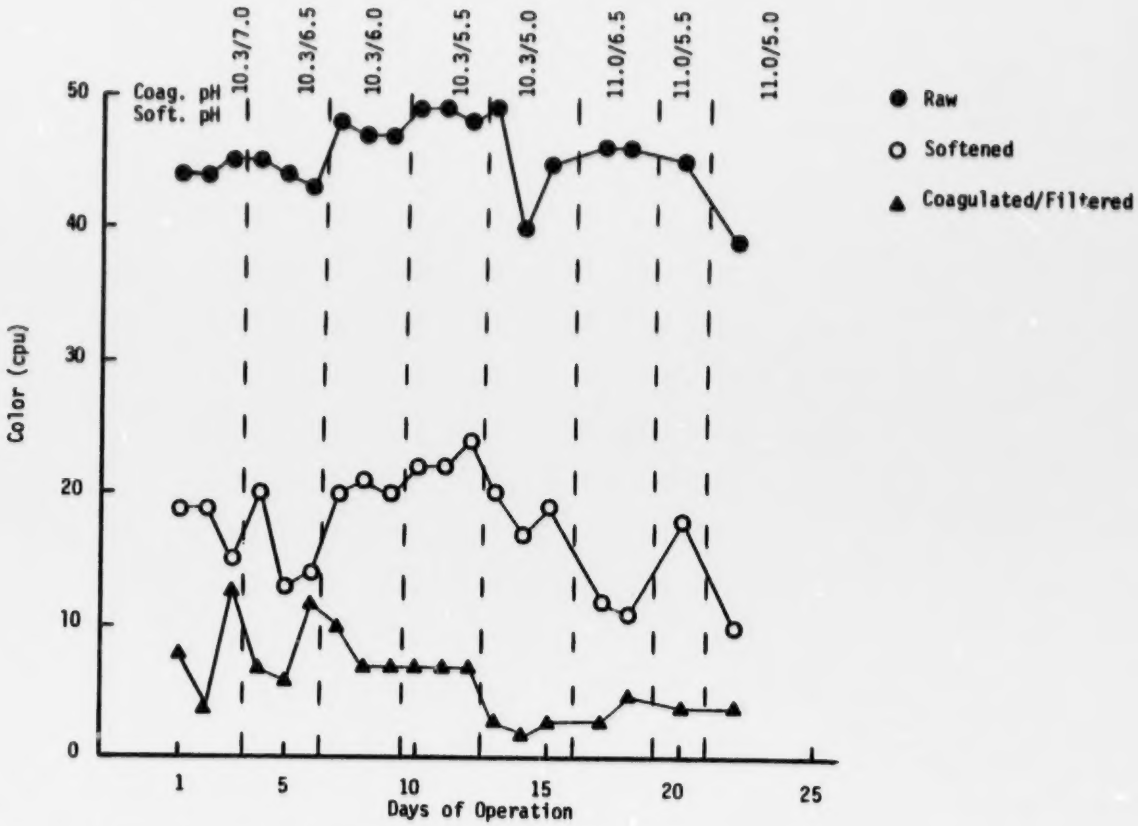


Figure 78 Raw, softened and coagulated color during Sequential Treatment Plant Test.

6.3.4.5 Hardness

The resulting hardness was approximately 120-mg/L as $CaCO_3$ or less when lime softening at pH 10.3 or less. As the softening pH increased above 10.3, the resulting hardness also increased due to the calcium added in excess of the raw water alkalinity. At pH 11.0, the total hardness exceeded 200-mg/L on one day as a result of the excess lime added. The excess calcium would normally be removed by recarbonation following softening, so these high values for hardness would not normally be experienced by excess lime softening.

6.3.4.6 DOC:THM:TOX Ratios

The DOC:THM:TOX ratios for the sequential treatment jar and plant tests are presented in Tables 63 and 64. The jar test DOC:THM:TOX ratios varied from 1:36.4:93.4 to 1:22.1:53.7 averaging 1:30:70 for coagulated water which had been softened at pH 10.3. When the softening pH was 11 this ratio averaged 1:20:71 varying from 1:18.1:57.9 to 1:22.9:81.3. The plant test samples averaged 1:47:00 overall varying from 1:80:144 to 1:37.2:80.4. The larger plant variations were probably due to a more difficult process control (pH). The THM:TOX ratios were 2.8 and 3.8 for the alum jar test on the pH 10.3 and 11.0 softening waters which were slightly higher than the 2.4 plant ratio.

6.3.4.7 Cost Summary

Table 65 summarizes the projected chemical cost for sequential operation at AID. These costs were calculated based on the doses applied during the plant trials and the unit cost (delivered) of each of the chemicals used.

Chemical costs for lime softening at pH 10.3 without alum coagulation were approximately $0.10 per thousand gallons. Alum coagulation increases chemical costs by adding the cost of alum and soda ash, along with costs for additional polymer and in some instances acid. Increases in cost ranged from 61 to 143 percent of the cost for standard lime softening. Best performance for THM reduction was achieved with an alum dose of 76 to 83-mg/L and a coagulation pH 6.0-6.5. The chemical cost of this treatment level was between $0.18 and $0.20 per thousand gallons. The average AID plant cost was $0.35/1000 gal. The percent increase chemical cost due to sequential treatment would increase the plant cost by 30 to 43%.

6.3.4.8 Summary

Lime softening followed by alum coagulation, was investigated on a bench and pilot plant scale at the Acme Improvement District water treatment plant located near West Palm Beach, Florida. Jar tests were conducted to determine the best sequence of softening and coagulation for THM precursor removal. Pilot plant testing was executed in two unused reactors that were serially connected for the project. There was no turbidity in the raw water. The hardness and alkalinity were typically 350-mg/L as $CaCO_3$ with color and DOC values of approximately 40 cpu and 15-mg/L as C.

TABLE 63. DOC/THM/TOX Ratios for Lime Softened and Alum Coagulated Water During Aid Sequential Treatment Jar Tests

Coagulation pH	Alum Dose* (mg/L)	Softening pH = 10.3		Softening pH = 11.0	
		DOC/THM/TOX	TOX/THM	DOC/THM/TOX	TOX/THM
5.0	29	1/22.1/53.7	2.43	1/16.3/77.1	4.74
5.5		1/18.6/69.6	3.75	1/16.4/66.0	4.04
6.0		1/19.2/72.9	3.80	1/13.7/81.1	5.91
6.5		1/28.3/72.2	2.55	1/25.4/75.1	2.96
7.0		1/26.2/61.7	2.35	1/19.3/66.0	3.43
5.0	58	1/21.7/60.9	2.80	1/18.0/65.5	3.63
5.5		1/20.2/80.5	3.98	1/10.3/70.0	6.77
6.0		1/22.0/70.4	3.21	1/23.1/69.5	3.01
6.5		1/41.5/69.4	1.67	1/27.5/74.0	2.69
7.0		1/38.9/78.1	2.01	1/25.9/76.6	2.96
5.0	117	1/30.8/80.3	2.61	1/18.1/69.4	3.83
5.5		1/68.4/20.9	3.27	1/14.2/71.1	5.01
6.0		1/18.1/68.8	3.79	1/23.1/67.7	2.93
6.5		1/18.4/73.7	4.01	1/17.8/69.6	3.91
7.0		1/33.4/74.9	2.24	1/23.7/82.1	3.46
5.0	175	1/36.4/93.4	2.57	1/18.1/57.9	3.19
5.5		1/26.8/61.5	2.30	1/16.3/60.3	3.69
6.0		1/27.3/65.7	2.41	1/15.7/69.4	4.41
6.5		1/25.3/67/6	2.67	1/23.0/67.9	2.95
7.0		1/34.0/68.4	2.01	1/27.9/81.3	2.91
Average		1/30/70	2.82	1/20/71	3.82
Lime Softened		1/18/54	3.0	1/19/55	2.9
Raw Water	DOC/THM/TOX	1/34/93	2.7		

*mg/L $Al_2(SO_4)_3 .14H_2O$

TABLE 64. Sequential Plant Treatment Tests

	DOC/THM/TOX	TOX/THM
3/29	1/86.1/92.9	2.57
3/30	1/32.3/139	4.30
3/31	1/31.4/84.6	2.69
4/5	1/32.5/90.5	2.78
4/6	1/20.4/103.5	5.07
4/6	1/27.3/83.4	3.05
4/13	1/35.0/81.9	2.34
4/13	1/49.9/94.6	1.89
4/14	1/35.5/19.7	1.96
4/14	1/37.2/10.4	2.16
4/15	1/45.3/96.4	2.13
4/15	1/50.4/84.5	1.68
4/16	1/68.9/112.3	1.63
4/20	1/79.3/116.3	1.47
4/21	1/80.0/144	1.80
5/3	1/53.6/109.2	2.04
5/4	1/59.4/124.3	2.09
5/5	1/68.1/127.8	1.88
5/5	1/87.2/135.3	1.55
Avg	1/47/100	2.4

TABLE 65. Sequential Treatment Plant Tests Cost Summary ($/1000 gal)

pH (lime softening)			10.3	10.3	10.3	10.3	10.3	11.0	11.0	11.0
pH (alum coagulation)			7.0	6.5	6.0	5.5	5.0	6.5	5.5	5.0
Lime Dose	mg/L	214	214	214	214	214	214	258	258	258
Alum Dose	mg/L	---	68	76	83	94	109	109	109	109
Acid Dose	mg/L	---	---	---	---	---	---	20	30	30
Soda Ash Dose	mg/L	---	10	20	30	30	30	30	30	30
Olin 5011 Dose	mg/L	0.4	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8
Chlorine		18	10	10	10	10	10	10	10	10
	$/lb				$/1000 Gal.					
Lime Cost	0.0367	0.0655	0.0655	0.0655	0.9655	0.0655	0.0655	0.0790	0.0709	0.0790
Alum Cost	0.0740	---	0.0420	0.0469	0.0512	0.0580	0.0673	0.0673	0.0673	0.0673
Acid Cost	0.0850	---	---	---	---	---	---	0.0142	0.0213	0.0213
Soda Ash Cost	0.1400	---	010117	0.0234	0.0350	0.0350	0.0350	0.0350	0.0350	0.0350
Olin 5011 Coast	4.9500	0.0165	0.0117	0.0330	0.0330	0.0330	0.0330	0.0330	0.0330	0.0330
Chlorine	0.1300	0.0195	0.0108	0.0108	0.0108	0.0108	0.0108	0.0108	0.0108	0.0108
		0.1015	0.1630	0.1796	0.1955	0.2023	0.2116	0.2393	0.2464	0.2464
% Increase			61	76	93	99	108	136	143	143

The initial bench scale sequential treatment work indicated that 60% of the DOC could be removed by sequentially treating AID raw water with softening and coagulation regardless of order. However, if pH was adjusted to 5.5 during coagulation, this DOC removal was increased 25%. Color removal was 85 to 90% and varied directly with DOC removal and THM precursor removal. THMFP was reduced to 167 to 217 μg/L by softening at pH 11.0 and coagulation at pH 5.5, which represented an increased average reduction of 45% relative to softening only at pH 10.3. TOX formation was generally three times THM formation.

The sequential treatment isopleths showed that softening at pH 11.0 averaged 41% DOC, 67% THMFP, 65% TOXFP and 63% color removal. This increased the removal relative to softening at pH 10.3 for each parameter (+11% DOC, +3% THMFP, +8% TOXFP, +3% color). Following lime softening with alum coagulation also increased organic removal. The optimum pH and dose were 5.0 to 6.0 and 117-mg/L alum. The DOC, THMFP, TOXFP and color were reduced an additional 30%, 13%, 23% and 30% respectively. Alum coagulation following softening at pH 10.3 increased the percentage removal of all the organic parameters more than alum coagulation following softening at pH 11. However the lowest concentrations of DOC, THMFP, TOXFP and color were from samples softened at pH 11.0 and coagulated at pH 5.0 to 6.0 although these differences were typically less than 5%.

Plant testing of sequential treatment was conducted by softening at pH 10.3 and 11.0 followed by alum coagulation from pH 7.0 to 5.0. The alum dose was varied from 60 to 109-mg/L. Figure 79 shows the average increase in DOC, THMFP, TOXFP, and color removal following softening at pH 11 or 10.3. The results show, as did the jar tests, that alum coagulation following softening will decrease the DOC, THMFP, TOXFP, and color relative to softening alone. The maximum DOC and color removals obtained in the plants tests were approximately 70% for DOC and 93% for color. The THMFP and TOXFP were reduced from 20 to 25%. There was slightly better removal for all parameters when pH 11 softening preceded coagulation as opposed to pH 10.3 softening as shown in Figure 79. The organic removal achieved in sequential treatment indicates that DOC, THMFP, and TOXFP can be reduced an additional 20-25% when alum coagulation is utilized in series with softening. The chemical cost increase would vary from $0.10 to $0.15 per 1000 gallons which would increase AID O&M cost ($0.35/1000 gal) by 30% to 43%. This data does indicate that, relative to the coagulation testing at other sites, adding softening to a coagulation process could not increase organic removal.

6.4 Membranes

6.4.1 Introduction

In the search for methods to control THM's in potable water, three generic means are apparent. The first is to remove the THM precursors (humic and fulvic acids) from the source water and the second is to remove the THM species once they have formed. The third method is the use of alternate disinfectants so that the necessary chemical reactions to form

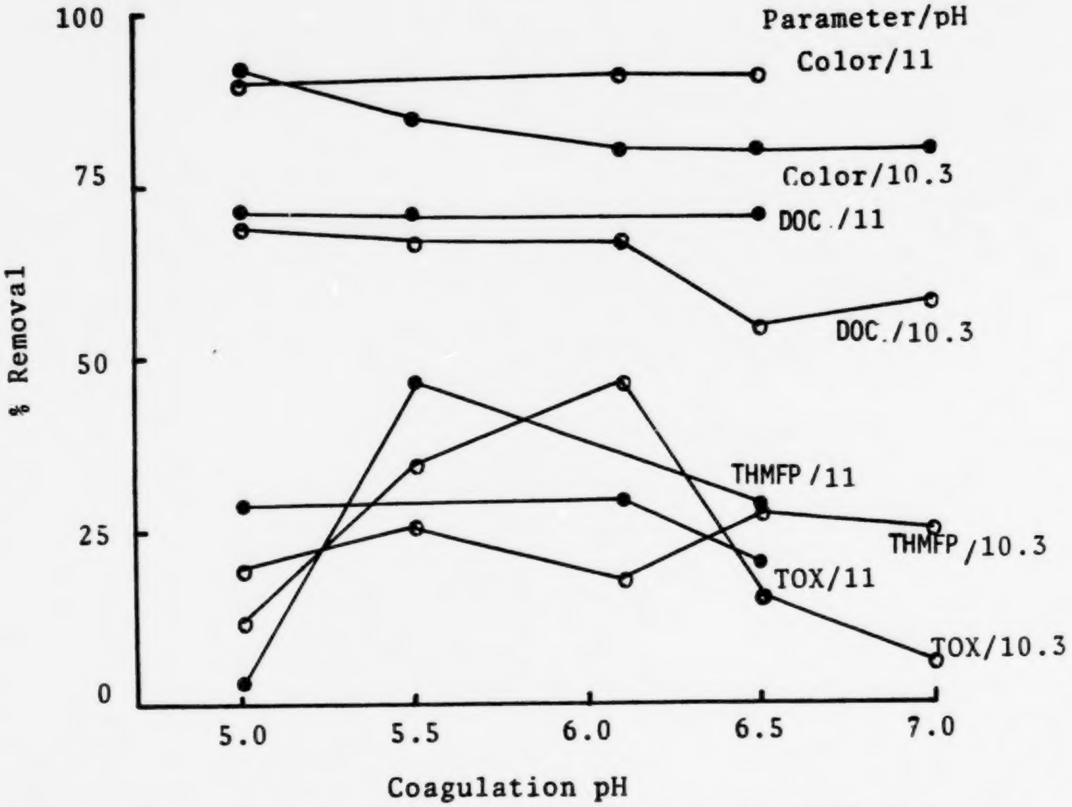


Figure 79 Average increase in percent DOC, THMFF, TOX and color removal by coagulation at varying pH for softened samples at pH 10.3 and 11.0. DOC and color based on raw water. THMFP and TOX based on softened water.

THM's cannot occur. To date no one method has been shown to be universally applicable or desirable.

The use of membranes for THM control works via the precursor removal mechanism. Depending on the size of the precursor and the retentive properties of the membrane material, different degrees of removal can be achieved. Generally the smaller the particle to be retained, the higher the energy costs due to osmotic pressure. The goal of this research phase was to identify a membrane that could suitably reduce the organic precursors in the feed water to a level that would allow the THM MCL to be met in the chlorinated product water. Secondary goals included the minimization of the energy requirements and the maximization of desirable potable water characteristics.

To achieve the stated goals, bench-scale testing was conducted on a number of membranes with various molecular weight cutoffs at the various sites. Following this analysis, an optimum membrane was chosen for pilot plant operations. In the pilot plant phase, the effects of varying feed pressure and product water recovery rate was studied along with the operation of the unit on a long-term basis. From the long term operation, changes in flux and product water quality were assessed as well as methods and effects of cleaning. At the Olga site the bench-scale unit was utilized to gain the same information albeit on a smaller scale.

The data gained from these studies is discussed in the following sections. Each of the testing sites is discussed separately and includes both bench-scale and pilot plant data. Following the individual site data, cost estimates for full-sized installations are given. Finally, a summary of the suitability and desirability of the process will be given.

6.4.2 Olga Membrane Selection Study

The initial step of membrane studies at the Olga water treatment plant was selection of the optimum membrane, which could "filter" the Caloosahatchee River water to meet the MCL for trihalomethane formation. A secondary objective was to determine product water values for inorganic parameters (Alk, TH, CaH, TDS, Cl, Na). Seven membranes with molecular cutoff weights ranging from 100 to 40000, were tested. The membranes tested, and their normal molecular weight cutoffs are listed in Table 66. Testing was initiated with the UFP 4040 membrane on October 29, 1984. Each membrane was operated approximately 24 hours to allow for river water acclimation, prior to sampling. Pretreatment consisted of prefiltration through a 5.0 micron filter. Duplicate samples were collected by Lee County personnel for concurrent analysis of inorganic parameters by UCF analysts and the Lee County laboratory. Organic analysis for DOC, THM's, TOXFP, and true color was conducted at UCF due to equipment limitations at Lee County. Lee County personnel were also responsible for recording pump feed pressure, product flow, and reject flows, as well as replacement and storage of membranes. The N-50 was the final membrane tested with operation completed on January 18, 1985. Numerous delays occurred during the testing sequence due to the following problems:

Table 66. Membrane Molecular Weight Rejection Capacity

Membrane		Molecular Weight Rejected
Filmtec	BW 30	100 MW
Filmtec	N-50	400 MW
Desal	U90-G10	2000 MW
Filmtec	UFP 4040	10000 MW
Desal	U90-G50	20000 MW
Osmotics	Osmo PT-2	20000 MW
Osmotics	Osmo PT-4	40000 MW

a) Broken inlet endpiece on membrane holder tube, requiring machining by Basic Technologies. 11/2 - 11/26.

b) 5.0 micron prefilter clogging - approximately every 10-12 hours of operation, causing unit to shut off.

c) Membrane unit was shut off when raw water service pumps were turned off, and was not always restarted when plant resumed production. Three days were needed to get 24 hours of membrane operation.

d) Christmas and New Years holiday periods, when insufficient personnel qualified to operate the unit were available.

Each membrane was run at a different feed pressure, as shown in Table 66, due to the wide range of possible flows (1.0 - 2.1 gpm), depending on molecular weight cutoffs. The product flow meter and reject flow meters had ranges of 0-2 gpm and 0-1 gpm, respectively. Percent recoveries for membranes with MW cutoffs greater than 1000 averaged 74% ranging from 66% to 81%. Percent recoveries for the NF-50 and BW 3030 were 50% and 33% respectively, due to low product flows. Hydraulic data based on flow meter readings was presented in Table 67. Generally, product flux increased as the membrane MW cutoff increased or as the operating pressure increased.

Table 68 shows THM's, TOXFP, and chlorine residual data for all seven membranes. Permeate DOC values are presented in Table 69. Based on product water DOC, initial chlorine does were estimated to ensure 48 and 96 hour residuals. The N-50 and BW 3030 membranes were the only membranes with 48 hour THM's less than 160 µg/L. The N-50 membrane met the 0.1-mg/L MCL, having a 48 hr THM's of 64 and 84 µg/L at Cl_2 dosages of 5 and 10-mg/L respectively. The BW 3030 48 hour THM's was 140 and 155 µg/L at the same chlorine dosages respectively, even though product water DOC for both membranes was 1.2-mg/L. The difference may be due to chlorine dosage variations, although 48 hr Cl_2 residuals for both membranes were 2.9-mg/L. Ninety-six hr. THM's for both membranes exceeded the 0.1-mg/L MCL at Cl_2 dosages of 5 and 10-mg/L. The 96 hr THM's for the N-50 was 102 µg/L essentially meeting the MCL. The 96 hr chlorine residuals were greater than 1.0 mg Cl_2/L at the 5-mg/L dose, indicating that dosage could be reduced. Operating pressures were 65 psi and 190 psi for the N-50 and BW 3030 respectively, making the N-50 the optimal membrane for organics rejection. It should also be noted that large difference in product water DOC and THM's existed for the U90-G50 and PT-2 membranes, both stated to have a MW cutoff of 20000.

Figure 80 shows percent removal of DOC and true color based on raw average values of 20.7-mg/L DOC and 73 cpu. The trend of decreasing values with decreasing M.W. cutoffs holds true except for the UFP 4040 membrane having a stated cutoff of 10000 MW. This MW was confirmed by the manufacturer, Filmtec, as being correct so the reasons for poor rejection of DOC and color are unclear. The DOC and true color values for the DeSal U90-650 and Osmo Pt-2 were averaged to yield one graph point for the 20000 MW. The plot of percent DOC removed (Figure 80b) indicates that a MW cutoff of 500 must be used before the THM MCL is met. Fifty percent of the molecules (by

Table 67. Olga Membrane Selection Study - Hydraulic Parameters

Membrane	Start Date	Molecular Weight Cutoff (MW)	Hours of Operation (Hr)	Feed Pressure (psi)	Feed Flow (gpm)	Product Flow (gpm)	Reject Flow (gpm)
DESAL U90-G10	Nov. 17, 1984	2000	20.0	100	1.7	1.2	.5
DESAL U90-G50	Nov. 26, 1984	20000	30.0	75	1.5	1.0	.5
OSMONTICS PT-2(411PS)	Nov. 30, 1984	20000	24.0	50	2.0	1.5	.5
OSMONTICS PT-4 (411TPS)	Dec. 19, 1984	40000	24.0	50	2.7+	2.2^+	.5
FILMTEC BW 30	Jan 2, 1985	100	24.0	190	1.5	.5	1.0
FILMTEC N50	Nov. 30, 1984	400	33.0	65	1.0	.5	.5
FILMTEC UFP 4040	Dec. 4, 1984	10000	24.0	55	2.1	1.6	.5

Table 68. Olga Membrane Selection Study TOXFP, THM's, and Chlorine Residual Data (48 and 96 hr)

Membrane	MWC	Cl_2 Dose (mg/L as Cl_2)	TOXFP (µg/L as Cl)		THM's (µg/L)		Cl Residual (mg/L as Cl_2)	
			48 Hour	96 Hour	48 Hour	96 Hour	48 Hour	96 Hour
U90-G10	2,000	5	574	625	274	326	NA	0.0
U90-G10	2,000	10	872	855	865	950	NA	1.6
U90-G50	20,000	10	1066	1367	510	500	0.0	0.0
U90-G50	20,000	20	1386	1794	690	900	4.0	2.0
PT-2 (411PS)	20,000	20	1954	2182	1165	1210	0.0	0.0
PT-2 (411PS)	20,000	30	2666	2616	1570	1810	2.7	0.0
N-50	400	5	122	186	64	102	2.9	1.6
N-50	400	10	139	206	84	131	8.0	7.2
PT-4 (411TPS)	40,000	30	2064	2098	1840	1840	8.4	2.5
PT-4 (411TPS)	40,000	40	2565	2887	2040	2280	10.3	8.6
UFP 4040	6,000	20	1840	1945	902	707	0.0	0.0
UFP 4040	6,000	30	3563	3655	1378	1050	3.7	0.0
BW 30 (R.O.)	100	5	191	201	140	140	2.9	1.4
BW 30 (R.O.)	100	10	195	250	155	200	5.3	5.2

Table 69. Olga Membrane Selection Study Raw Product Parameters and Removal Efficiencies

Membrane and Sample	MW Cutoff (mw)	DOC (mg/L as C) *	Color (cpu) *	Alk. (mg/L $CaCO_3$)	TH (mg/L $CaCO_3$)	CaH (mg/L $CaCO_3$)	TDS (mg/L)	Temp (°C)	pH	Turb. (NTU)	Cl^- (mg/L)	Na^+ (mg/L)
DESAL U90-G10	2000	7.08	12.5	102	190	180	402	28	7.5	0.4	87	41
Raw		15.93	70.0	110	220	204	484	28	7.9	5.0	90	43
% Removal		55.6	82.1	7.3	13.6	11.8	16.9	--	--	--	3	2
DESAL U90-G50	20000	10.70	14.0	132	204	200	358	23	8.1	2.3	60	29
Raw		23.71	80.0	137	214	200	384	23	8.0	0.2	60	29
% Removal		54.90	82.5	3.6	4.6	0.0	6.7	--	--	2.3	0	3
OSMO PT-2 (411Ps)	20000	17.50	67.0	135	228	200	372	22	7.8	1.0	73	38
Raw		22.04	75.0	140	234	220	424	22	8.2	0.3	75	38
% Removal		21.	11.0	3.6	2.5	9.1	12.3	--	--	--	2	0
OSMO PT- (411TPS)	40000	17.88	51.0	130	240	220	436	22	7.9	0.2	117	56
Raw		18.12	68.0	131	246	226	470	22	7.5	0.7	117	56
% Removal		1.3	25.0	3.1	2.4	2.6	7.2	--	--	--	0	0
FILMTEC BW30	100	1.19	2.5	12	16	12	56	19	7.8	0.08	16	9
Raw		21.84	67.0	143	252	208	605	19	7.5	0.8	175	96
% Removal		94.6	96.2	92.3	93.6	94.3	90.7	--	--	--	91	91
FILMTEC N50	400	1.24	3.0	31	46	36	86	22	7.6	0.2	30	17
Raw		21.80	70.0	127	202	180	334	22	7.5	0.7	63	30
% Removal		94.3	95.7	75.6	77.2	80.0	83.5	--	--	--	52	57
**FILMTEC UFP 4040	10000	18.96	70.0	127	194	186	332	21	8.1	.3	71	35
Raw		21.59	80.0	130	206	190	336	21	7.5	.6	71	36
% Removal		12.2	12.5	2.3	5.8	2.1	1.2	--	--	--	6	3

* UCF Data - All other parameters Lee County Lab Data except UFP 4040**

** All UFP 4040 Data is UCF Data.

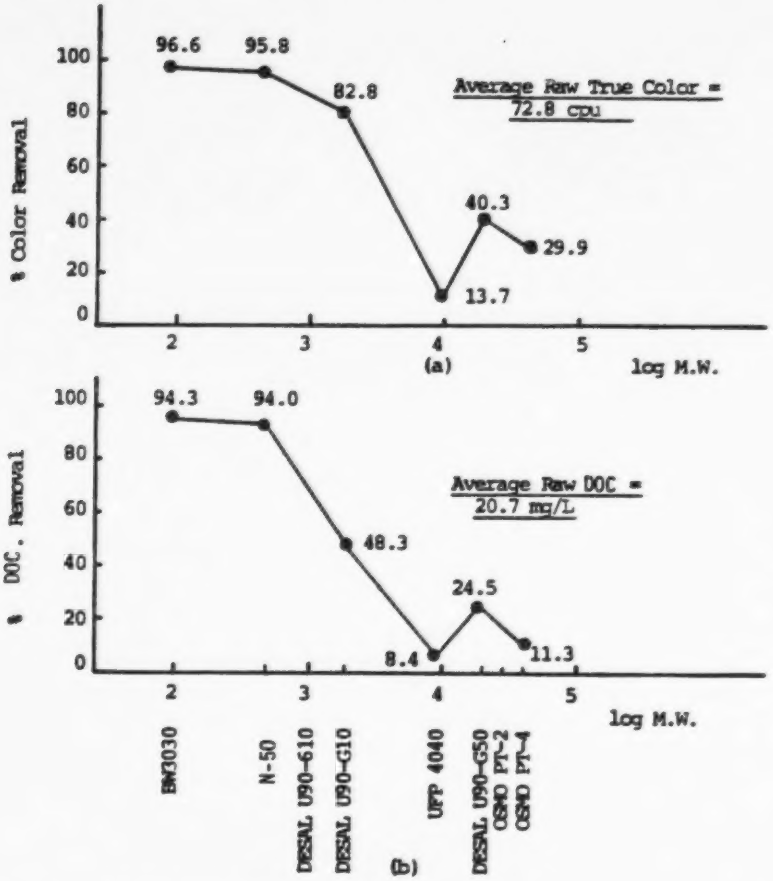


Figure 80 (a) Percent removal of true color and (b) percent removal of DOC. As a function of molecular weight size fractionization using membranes with different molecular weight cutoffs, Olga, Florida.

mass) causing DOC lie in the 500-2000 MW range based on DOC removed. An additional 25% have MW's between 2000 and 20000.

The plot of percent true color removal (Figure 80a) follows the same pattern as the percent DOC plot. This indicates that true color is organic in nature. Four percent (by mass) of the molecules causing true color have MW's less than about 1000 MW, 13% lie in the 500 to 2000 MW range, and 83% have MW's greater than about 2000 MW. A color of 12.5 cpu was observed at 2000 MW, roughly indicating that a MW of 2000 is necessary to meet the 15 cpu standard. The product water from the Osmo PT-2 (MW = 20000) and UFP 4040 (MW = 10000) had a true color of 67 and 70 cpu respectively.

Table 69 lists the inorganic parameter data for all seven membranes. DOC and color values are also presented. Figure 81 shows percent removal of chlorides, total hardness and alkalinity as a function of membrane MW cutoffs. Generally, total alkalinity, TH, CaH, Cl and Na permeate concentrations decreased with decreasing molecular weight cutoffs.

Permeate alkalinities declined as molecular weight cutoffs decreased. Figure 81c indicates that 22% (by mass) of the molecules comprising total alkalinity were larger than 2000 MW, 54% were in the 500 to 2000 MW range and 24% had MW less than about 500. The N-50 and BW 3030 had total alkalinities of 31 and 12-mg/L respectively. Although manufacturer's molecular cutoff weights are approximations, size fractionalization using several different MW cutoff membranes allows the molecular size of inorganics or organics removed from the water to be determined.

Product total hardness varied from 240-mg/L to 16-mg/L $CaCO_3$ as MW cutoffs decreased from 40000 to 100. A MW cutoff of 500 was necessary to achieve a permeate total hardness of 100 mg $CaCO_3$ or less. Figure 81b indicates that 80% of the constituents comprising total hardness have MW greater than 500, and 16% have MW greater than 2000. The N-50 and BW 3030 permeates had TH values of 46 and 16-mg/L $CaCO_3$ respectively.

Calcium hardness (CaH) rejection followed the same trend as TH. An 80% reduction (by mass) of CaH required membranes having MW cutoffs less than 2000. N-50 and BW 3030 CaH was 36 and 12-mg/L $CaCO_3$ respectively.

Significant permeate chloride rejection did not occur until MW cutoffs were 2000 or less as shown by Figure 81a. The U90-G10 membrane (MW = 10000) rejected 3% of the raw concentration. At MW cutoffs of 500 and 100; Cl rejections were 52 and 94 percent respectively. Chloride ion molecular weight is less than 2000.

Sodium rejection did not significantly occur until MW cutoffs of 500 or less were used. Rejections were 57 and 91% (by mass) for the N-50 and BW 3030 membranes, respectively.

The N-50 membrane permeate had a pH of 8.1, total alkalinity of 31, and CaH of 36-mg/L $CaCO_3$, yielding a pH's of 8.6, which is slightly corrosive. In comparison, BW 3030 permeate had a pH of 7.5, total alkalinity of 12 and a CaH of 12-mg/L $CaCO_3$, resulting in a pH's of 9.6, which is highly

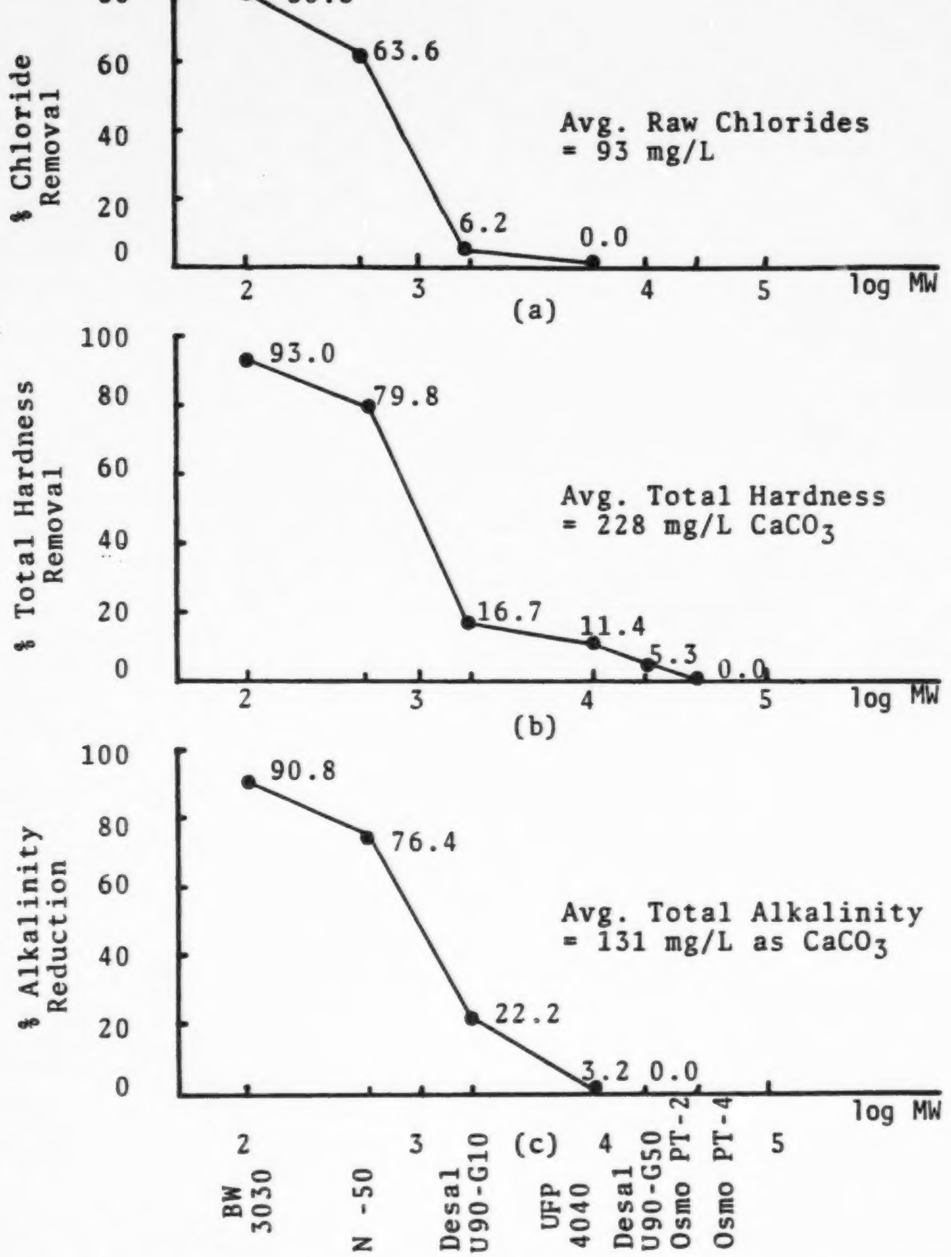


Figure 81 Percent removal of (a) chlorides (b) total hardness, and (c) alkalinity , as a function of molecular weight-size fractionization using membranes with different M.W. cutoffs, Olga, Florida.

corrosive. Thus, the N-50 membrane produced a nearly stable water with an acceptable hardness, making it the optimum membrane of those tested for inorganic removal. The normal feed pressure range for the N-50, is 50 to 120 psi, giving it a considerable advantage over the 150 psi - 300 psi range required for the BW 3030 R.O. membrane.

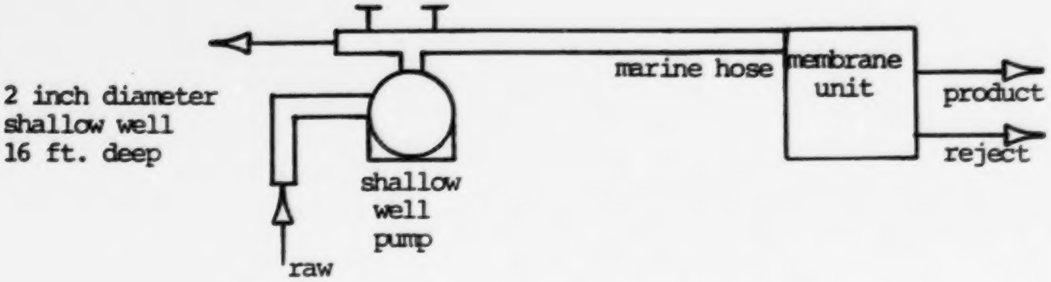
It is true that a lower percentage of recovery would improve permeate water quality and possibly a low enough recovery could be utilized so that the permeate would meet any MCL. However UF has been noted to act like a molecular-sieve and not exhibit the adverse effects of an increasing feed concentration on permeate water quality, but this may not be the case in potable water. In the interest of time and with consideration of an UF controlled process, it was decided that the Filmtec N-50 membrane was the optimal choice for continued evaluation. This choice was due to the N-50's excellent organic and inorganic rejection properties. The ability to operate at low feed pressures, favors its use over conventional RO membranes. The next step in the evaluation process was a continuous long term study, to determine feed pressure and percentage of recovery conditions which maximize organic and inorganic rejections.

6.4.3 Membrane Operating System

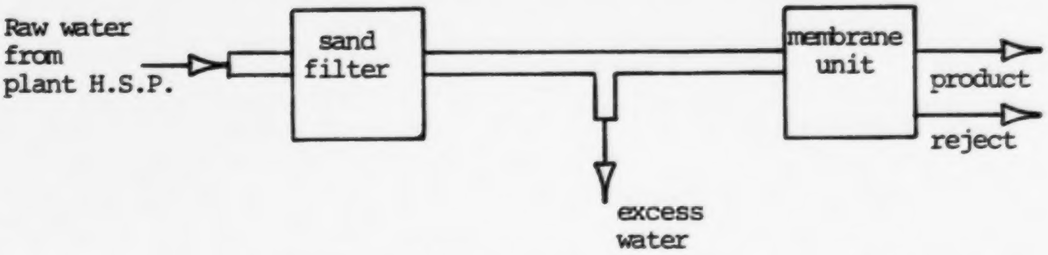
Ultrafiltration requires a "feed" water which is low in turbidity (less than 1.0 NTU). The "007" unit was designed with a cloth fiber filter cartridge to remove turbidity and suspended solids prior to membrane filtration. During the optimum membrane selection testing, the 5.0 micron prefilter became clogged after 10 to 12 hours of operation, causing the membrane unit to automatically turn off due to low feed pressure. Additionally, alum was fed in the raw water intake channel and may have clogged 5.0 micron prefilter. To lessen these problems over long periods of operation, a physical pretreatment system was needed. Four configurations as shown in Figure 82 were tried before a satisfactory working system evolved. The optimum operating system was in place on March 13, 1985.

The first system (a) used a shallow well to "prefilter" river water through the river bank. A shallow well pump (Goulds model XSH10) was selected to deliver a flow of 10 gpm at a feed pressure of 40 psi to the membrane unit. The shallow well was only able to produce 1.5 gpm due to the high organic content of the river sediments. Consequently, insufficient feed pressure was available to the membrane unit, and flows were inadequate for membrane "flushing" during normal operation.

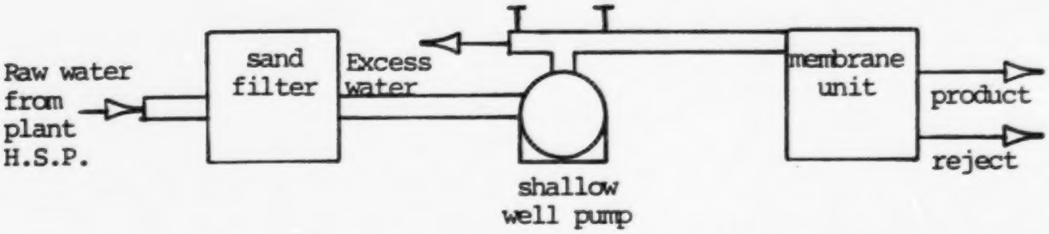
The second configuration (b) used a Diaclear brand silicia sand filter (MS-16), with backwash capability. The sand had a nominal diameter of 0.45 to 0.55 mm and a uniformity coefficient of 1.6. Raw water was supplied from a hose bib on the water plant raw water pump. To provide feed pressure to the membrane unit, about 10 gpm of raw water had to be filtered. Adequate feed pressure was provided until the sand filter began to clog. As filter backpressure rose, reduced output pressure caused membrane unit shutdown. Sand filter backwashing was needed about every 6-8 hours due to the high filtration rate (10 gpm/ft^2). The sand filter reduced turbidity from 4 to 1 NTU or less and had no effect on DOC.


2 inch diameter
shallow well
16 ft. deep
marine hose
membrane unit
product
reject
shallow well pump
raw

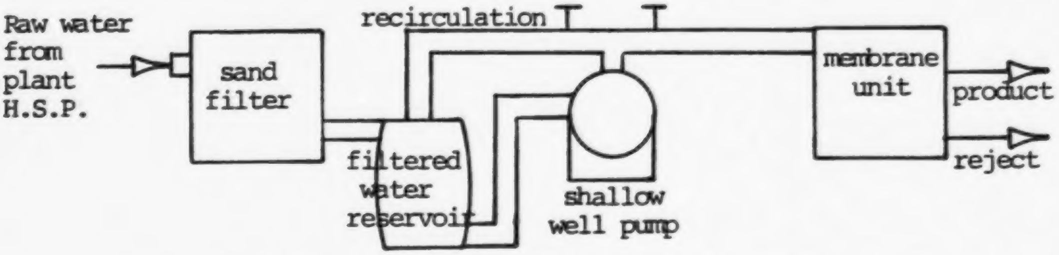
(a) Shallow Well with Direct Feed to Membrane Unit.


Raw water from plant H.S.P.
sand filter
membrane unit
product
reject
excess water

(b) Addition of Sand Filter Using Plant Raw Service Pump.


Raw water from plant H.S.P.
sand filter
Excess water
membrane unit
product
reject
shallow well pump

(c) Addition of Shallow Well Pump to Provide Feed Pressure.


Raw water from plant H.S.P.
sand filter
recirculation
membrane unit
product
reject
filtered water reservoir
shallow well pump

(d) Addition of Reservoir with Recirculation Loop.

Figure 82 Flow schematics showing evolution of operating system for bench scale N -50 membrane unit, Olga, Florida.

The third configuration (c) added a shallow well pump to boost feed pressure to the membrane unit. Flows were reduced to 5 gpm to meet the shallow-well, pump-head, capacity curve limits, and reduce sand-filter clogging. This system operated well, if the sand filter was backwashed frequently.

The final configuration (d) added a filtered-water reservoir between the sand filter and the shallow well pump, allowing the sand filter filtration rate to be reduced to less than 1.0 gpm during routine operation. Backwashing intervals were greatly increased (2 to 3 days). Recirculation of the filtered water through the reservoir allowed the shallow well pump to be operated efficiently. It should be noted that the membrane system was shut down coincident with the raw water service pumps. The membrane system did not operate from March 2 to March 13 due to the three system modifications, and from March 14 to March 28 due to direct alum feed into the raw water pump intakes. The membrane unit, holding tank, and sand filter for the extended study are shown in Figures A-12, A-13, and A-14.

6.4.4 Operational Matrix

On February 28, 1985, after approximately 100 hours of N-50 operation, a test matrix was developed to determine the effects of feed pressure and product recovery on product quality. Previous testing during the membrane selection process had been done at 60 psi and a 50 percentage of recovery ratio. The matrix feed pressure was varied from 60 to 120 psi at product recoveries of 50, 60, 70, and 80% each. Each experimental setting was operated for 30 minutes to allow for stabilization. At the end of the period, product water samples were collected for analyses of THM's, DOC, color, pH, Alkalinity, TH, Calcium hardness, TDS, chlorides and sodium. Reject samples were also collected to determine reject TH and chlorides for mass balances. The results are presented in Tables 70 and 71. Reductions in raw parameters include pretreatment as well as membrane filtration. Analysis of Table 71 indicates that DOC and THM generally decreased as feed pressure and product flow increased. THM's (dosage = 5 mg Cl_2/L) at 48 and 96 hours tended to increase as product recoveries increased at a constant pressure. The lowest THM's occurred at pressures of 105 psi and 120 psi for product recoveries of 50 to 70%. Color was relatively constant over the product recovery range at a constant feed pressure, and decreased as feed pressure increased. Ninety-six hour chlorine residuals greater than 1.0-mg/L were present under all conditions. Mass balances using reject values yielded errors less than 15% for TH and Cl at all conditions except 80 percentage of recovery. At 80% recovery, errors exceeded 25% indicating variability of reject flows due to low volume. Flux was determined directly by measuring flow volume over time and dividing by a known membrane surface area. Product fluxes ranged from 12.5 gallons/day ft^2 (640 mls/min) at 60 psi to 24.2 gallons/day ft^2 (1240 mls/min) at 120 psi.

Analysis of Table 71 indicates a narrow range of values for pH, alkalinity, total hardness, calcium hardness, TDS, chlorides and sodium over the entire range of feed pressures. Inorganic parameters tended to decrease with increasing recovery at a constant pressure. Alkalinity, total hardness, and calcium hardness averaged 44, 56, and 42-mg/L as $CaCO_3$,

Table 70. Olga Ultrafiltration Matrix, February 28, 1985 - ORGANICS*

Sample ID (psi)	Percent Recovery (%)	Dissolved Organic Carbon (mg/L as C)	48 Hr THM's Formation (µg/L)	96 Hr THM's Formation (mg/L as Cl_2)	96 Hr Cl_2 Residual (mg/L as CL_2)	Color (cpu)
RAW	----	19.6	0	0	----	63
60 psi,	50%	2.3	106	108	1.7	3.5
60 psi	60%	2.6	128	138	1.5	3.5
60 psi	70%	2.6	119	142	1.3	4.0
60 psi	80%	2.6	91	122	1.2	4.0
75 psi	50%	1.3	84	89	2.1	3.0
75 psi	60%	1.3	97	98	2.1	3.0
75 psi	70%	1.5	100	101	2.4	3.5
75 psi	80%	1.7	130	164	1.6	3.0
90 psi	50%	1.2	114	110	2.3	2.5
90 psi	60%	1.4	84	127	2.2	2.5
90 psi	70%	1.7	79	106	2.5	2.0
90 psi	80%	1.7	117	120	2.4	2.0
105 psi	50%	1.4	71	73	3.7	2.0
105 psi	60%	1.6	58	72	3.5	2.0
105 psi	70%	1.5	83	85	2.6	2.0
105 psi	80%	1.1	110	122	2.7	2.0
120 psi	50%	1.0	58	62	2.4	1.5
120 psi	60%	1.0	68	71	3.4	1.5
120 psi	70%	1.3	61	61	4.3	1.5
120 psi	80%	1.4	79	82	3.4	1.5

Table 71. Olga Ultrafiltration Matrix, February 28, 1985 - INORGANICS*

Feed Pressure (psi)	Percent Recov. (%)	pH	Alkalinity (mg/L $CaCO_3$)	Total Hardness (mg/L $CaCO_3$)	Calcium Hardness (mg/L $CaCO_3$)	Total Diss. Solids (mg/L)	Chlorides (mg/L)	Sodium (mg/L)
RAW	----	8.2	160	245	197	518	122	60
60	50	7.6	50	60	42	161	55	35
	60	7.5	52	68	50	189	65	38
	70	7.5	58	66	48	183	60	38
	80	7.5	64	65	50	157	57	33
75	50	7.6	40	54	32	142	62	44
	60	7.6	42	56	40	222	67	42
	70	7.7	49	56	43	171	77	47
	80	7.7	44	61	50	219	81	48
90	50	7.8	40	56	38	164	63	39
	60	7.8	43	58	42	174	68	42
	70	7.7	50	57	48	194	76	45
	80	7.7	52	63	55	149	76	47
105	50	7.7	32	49	34	142	63	36
	60	7.7	38	48	38	152	70	40
	70	7.7	42	54	42	178	69	45
	80	7.7	46	60	46	178	76	49
120	50	7.7	36	38	28	144	55	36
	60	7.7	40	44	32	177	63	37
	70	7.7	46	52	38	157	69	45
	80	7.7	48	58	48	186	74	47

respectively, at a pH of 7.7. Product TDS ranged from 142-222-mg/L, while chloride and sodium rejection averaged 45% and 35% respectively.

Another smaller test matrix was conducted on March 29, 1985, to verify the results of the initial matrix. Feed pressures were varied from 75 to 120 psi at product recoveries of 60 and 70%. Raw water quality with the exception of TDS and chlorides was nearly identical to the previous matrix. Analysis of Table 72 indicates product DOC and color values consistent with previous results. THM's (96 hr) averaged only 25 µg/L higher, while chlorine residuals were the same. A lower chlorine dose of 4-mg/L could be used to reduce the THM's and still maintain required residuals. The inorganic parameter product data presented in Table 73 is consistent with the previous matrix. Product alkalinity, total hardness and calcium hardness averaged 60, 52, and 41-mg/L as $CaCO_3$. Chloride rejection averaged 38% with a mean value of 62-mg/L. Flux rates during the second matrix were identical to the initial matrix. The membrane surface was relatively "clean" during the matrix testing as indicated by excellent organic removal.

Matrix data for organic and inorganic parameters was reformatted so that product water parameter values could be expressed as percent removals from the raw water values. Percent removals were plotted as function of feed pressure and percentage of recovery. Figures 83, 84, and 85 depict the removal efficiencies for THM's, DOC, and color respectively. Figures 86, 87, and 88 show removal efficiencies for alkalinity, total hardness, and chlorides, respectively. A brief discussion of the figures follows.

a) THMFP - The THMFP for raw Caloosahatchee river water was determined by dosing at 5-mg/L at pH 8.0 for 4 different samples. Ninety-six hr THM's were 480, 668, 428, and 555 µg/L resulting in a mean value of 533 µg/L. Figure 83 indicates that THM's meets the MCL at high pressures and low percent recoveries. All THMFP's were normally measured at pH 8.0, 25°C, 96 hrs, and generally 5-mg/L Cl_2 dose. Sometimes the Cl_2 dose was varied so the Cl_2 dose is specified for all THMFP data, and if the conditions are not stated then they were normal.

b) Total Organic Carbon - (Figure 84) DOC reduction exceeded 86% under all conditions. DOC removal of greater than 93% based on a raw DOC of 18.8-mg/L, was sufficient to meet the THM MCL.

c) Color removal - Figure 85 indicates that color removal exceeded 93% under all conditions based on a raw color of 63 cpu. Color decreased as feed pressure rose, but was relatively insensitive to percentage of recovery.

d) Alkalinity - (Figure 86) Percent reductions of at least 63% were attained for all operating conditions. Alkalinity reduction was relatively insensitive to feed pressure, but was affected by percentage of recovery.

e) Total Hardness - Figure 87 indicates at least 70% total hardness removals at all operating conditions, based on a average total hardness of 242-mg/L as $CaCO_3$.

Table 72. Olga N-50 Ultrafiltration Matrix March 29, 1985
Organic Parameters

Sample Type	Feed Pressure (psi)	Percent Recovery (%)	Dissolved Organic Carbon (mg/L as C)	96 HR THM's (μg/L)	Color (cpu)	96 HR Cl_2 Residual (mg/L as Cl_2)
Raw	———	———	18.2	668	55	0.0
Product	75	60	1.2	107	1.0	2.6
Product	90	60	1.5	100	1.0	2.6
Product	105	60	1.2	114	1.0	2.4
Product	120	60	1.3	100	1.0	3.4*
Product	75	70	1.5	120	1.5	2.1
Product	90	70	2.0	136	1.5	2.7
Product	105	70	1.5	124	1.0	2.8
Product	120	70	1.4	129	1.0	2.7

* Cl_2 overdose 7.0 mg/L instead of 5.0 mg/L.

Table 73. Olga N-50 Ultrafiltration Matrix March 29, 1985
Inorganic Parameters

Sample Type	Feed Pressure (psi)	Percent Recovery (%)	pH	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Calcium Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)	Chlorides (mg/L)
Raw	———	———	8.1	174	238	190	409	100
Product	75	60	7.7	60	54	48	83	62
Product	90	60	7.5	60	54	44	130	62
Product	105	60	7.4	52	54	36	58	57
Product	120	60	7.4	54	44	34	59	55
Product	75	70	7.8	58	48	48	76	65
Product	90	70	8.0	54	46	46	117	68
Product	105	70	7.9	50	38	38	91	62
Product	120	70	7.8	46	35	35	97	64

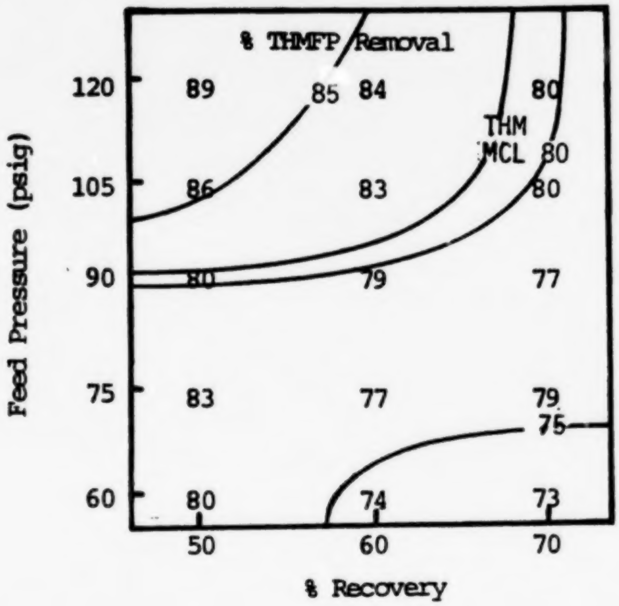


Figure 83 THMFP isopleth for N -50 based on 96 hr formation potential of 533 ug/l at 25°C for Feb. 28 and March 29 matrices.

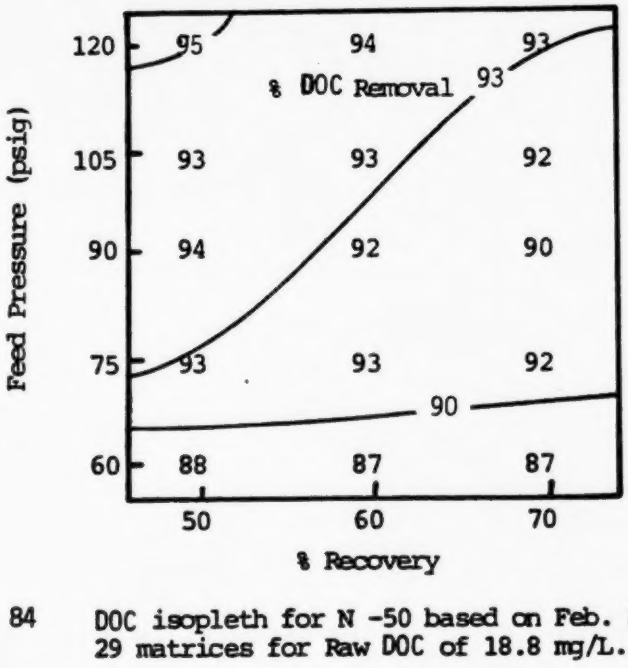


Figure 84 DOC isopleth for N -50 based on Feb. 28 and March 29 matrices for Raw DOC of 18.8 mg/L.

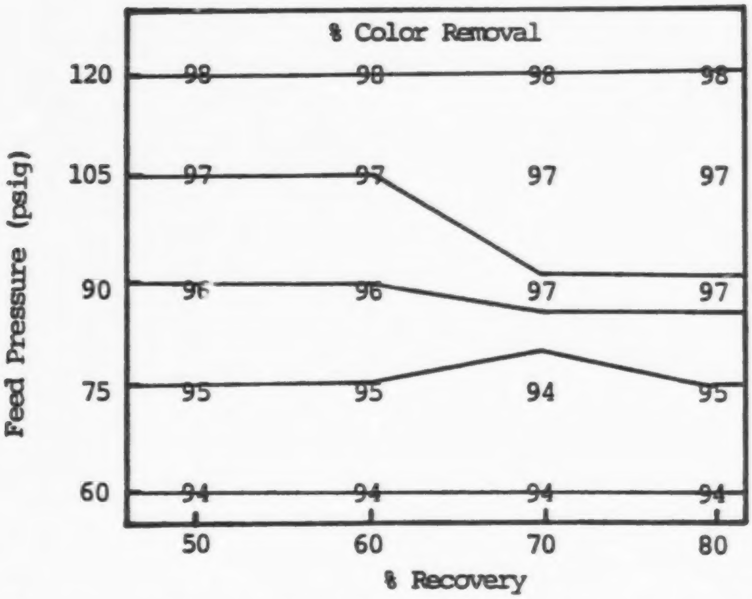


Figure 85 Product water true color isopleth for N -50 based on Feb. 28 matrix (Raw true color = 63 cpu).

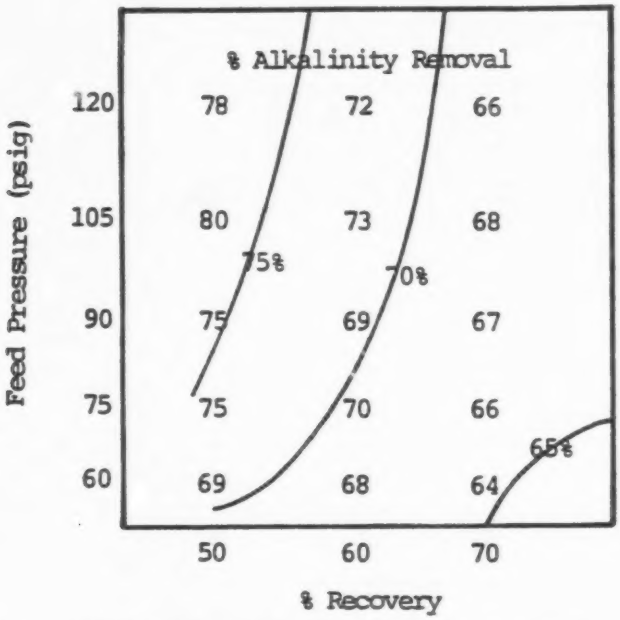


Figure 86 Alkalinity isopleth for Olga N -50 based on Feb. 28 and March 29 matrices (Raw alkalinity = 167 mg/L $CaCO_3$).

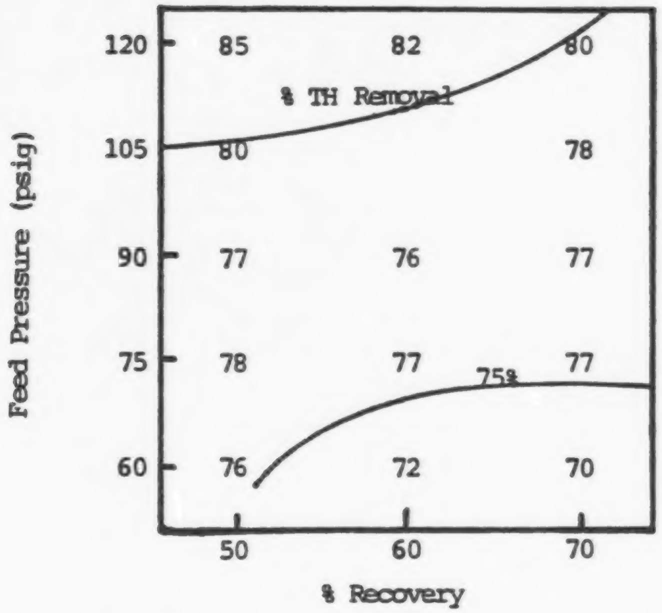


Figure 87 Total hardness isopleth for N -50 based on Feb. 28 and March 29 matrices. (Raw total hardness = 242 mg/L $CaCO_3$)

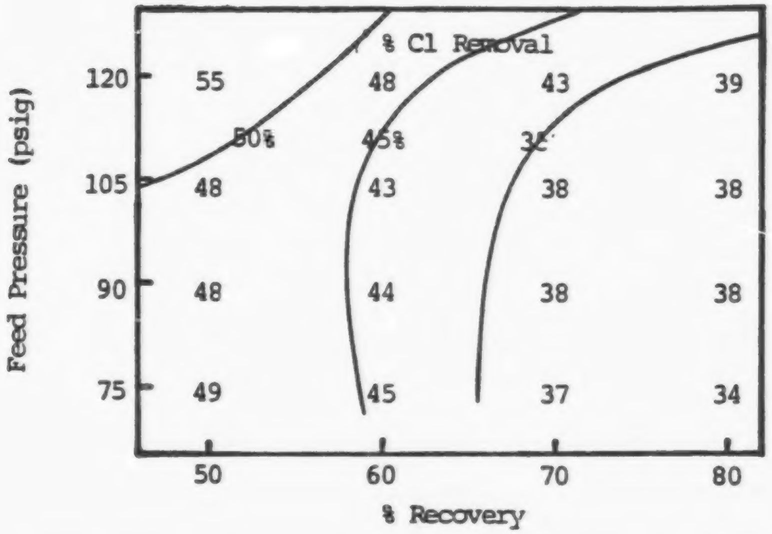


Figure 88 Chloride Isopleth for Olga N -50 based on Feb. 28 matrix. (Raw water Cl = 122 mg/L).

f) Chloride removal - Percent chloride removal was greatest for feed pressures greater than 105 psi, and at 50% recovery. Figure 88 indicates that chlorides significantly increase at 70 and 80% product recoveries.

6.4.5 Extended Study

The N-50 membrane was selected for additional study based on the product water quality produced during the optimum membrane testing matrix. This membrane rejects molecular weights of 400 or more and had a total surface area of 19.5 square feet. Testing was conducted between February 22 and May 17, 1985 using a bench scale unit which produced approximately 400 gpd. The objectives of the study were to determine:

(1) operating conditions (pressure and percentage of recovery) required to meet the THM MCL.

(2) product water values for DOC, TOX and color

(3) product water values for pH, TH, CaH, TDS, Cl, and Na.

(4) product water flux rates for varying feed pressure and product recovery

(5) effects of membrane cleaning on product quality and flux.

All sample laboratory analyses except sodium were performed by UCF personnel, either at UCF, or at the Olga water treatment plant lab. Samples were collected twice per week for a forty day period by either UCF or Lee County personnel, who were also responsible for routine operation of the membrane unit and recording of flux and pressure measurements. The long term testing consisted of two N-50 matrices, and an extended operation study.

N-50 MEMBRANE EXTENDED OPERATION STUDY

The second, and major part of the Olga N-50 UF study was initiated on April 11, 1985 to verify test matrix results under extended operation. Long duration testing was needed to quantify organic and inorganic removal efficiencies for variable raw water characteristics and membrane operating conditions. Permeate flux at feed pressures of 60 to 105 psi was monitored over time to account for flux decline due to membrane fouling. No chemical pretreatment of the raw water, such as pH adjustment or antiscalant, was used. Physical pretreatment consisted of sand filtration followed by a 5.0 micron cloth fiber filter. The membrane was chemically cleaned once to investigate flux restoration and its effect on product water quality. A maximum feed pressure of 105 psi was used in long term operation by manufacturer's recommendation, since the N-50 is structurally a "low pressure" UF membrane.

Sample collection was conducted by Lee County or UCF personnel every 3 to 4 days over the forty day test period. On two occasions, April 25 and May 6, 1985, two samples were collected by UCF researchers. A minimum of 8 hours operation occurred between sampling to allow for stabilization at different pressures and recovery ratios. Additional samples were taken before and after membrane cleaning (April 18, 1985). Samples were refrigerated at the Olga water treatment plant until transport to UCF. Plant operators recorded permeate flow, reject flow, and operating pressures about twice per shift. Conditions between readings were assumed constant, except as noted in the membrane operations log book.

Before discussing product water quality and percent removals as a function of operating conditions, the distinction between "raw" water and "feed" water should be made. "Raw" water has the same physical and chemical components as the Caloosahatchee River source. Percent removals based on "raw" use laboratory filtered (0.45 u glass fiber) raw water as a basis. Feed water consists of a blend of 5.0 u prefiltered-raw and recirculated "reject" water, which contains the rejected components of the product water. Feed water chemical and physical characteristics actually "contact" the membrane surface. Provision for feed water sampling was not provided in the "007" membrane unit used in this study. Consequently, feed-water quality concentrations were calculated by mass and flow balances using measured raw and reject water concentrations. Flow through the Procon brand feed pressure pump was constant at 4 gpm over the 60 psi to 105 psi range, based on the manufacturers pump curve. The recirculation flow rate was equal to the pump flow rate minus the sum of the product and reject flows. Feed concentrations were computed on a weighted flow basis. As product flow increased, the percentage of feed flow attributable to the raw flow increased, and the proportion of feed flow due to the recirculation stream was reduced. Thus, higher product flow rates yielded lower "feed" concentrations at the membrane surface. The long-term study was run at recirculation ratios greater than 88% due to low product flows (600-1200 ml/min) through the N-50 membrane. The recirculation control valve was fully open throughout the test duration, allowing scouring of the reject side of the membrane in order to prolong membrane life.

The reject flow depended on the desired percentage of recovery of the raw water. Percentage of recovery is defined as the ratio of the product water flow divided by the raw water input. If 0.6 liter/minute of product water is produced at 60% recovery, the reject flow is 0.4 liter/minute. Low product flows (600 ml/min) at low pressures yielded very low reject flows. The reject control valve tended to close under low flows, and had to be adjusted frequently.

Test results were evaluated on a raw water basis, since raw water values were analytically determined. Percent removals based on computed "feed" concentrations are specifically noted. The extended operation study is discussed in four sections; permeate flux, membrane cleaning, organic parameters, and inorganic parameters. A summary based on Olga test results, follows the discussion.

6.4.5.1 Flux

The operating condition of the membrane surface is a significant factor in permeate flux and quality. Membrane conditions varied from very clean near the beginning of the study (matrices) to partially fouled near the end of the study. Feed pressure and percentage of recovery are plotted against hours of operation, in Figure 89.

Product water flux always declined with time. (A steady state flux was never achieved.) For example, fluxes at 105 psi varied from an initial value of 18 gpd/ft^2 to 14.2 gpd/ft^2 after 150 hours of operation, a decline of 20%. Permeate quality remained almost constant during this interval. Permeate flux at 60 psi (hours 260-330) varied from 14.6 gpd/ft^2 to 12.8 gpd/ft^2 over 70 hours, a decline of 12%. Average permeate flux at 60 psi (hours 427-436) was 11.45 gpd/ft^2 indicating foulant layer deposition during the interim period of operation, when different operating conditions were used.

Permeate fluxes were relatively independent of percentage of recovery. Flux matrices run at 33 and 75% recovery showed only minor variations in flux, over the entire feed pressure range of 60 to 120 psi. Product fluxes at 75% recovery were slightly higher than at 33% recovery, increasing 0.59 gpd/ft^2.

Permeate flux at pressures of 60, 70, 75, 90, 95, 105, and 120 psi was statistically analyzed as shown in Table 74. Membrane cleaning occurred prior to the 95 psi setting. Average values for 60, 70, 75, 90, 105, and 120 psi yielded the following relationship by linear regression:

$$\text{Flux} = 0.071 \text{ (feed pressure)} + 8.11.$$

This equation ($r^2 = .868$) applies for pressures from 60 to 120 psi, and is plotted in Figure 90.

Large standard deviations, and ranges are attributable to variable membrane conditions or mechanical variations in the "007" unit operating system i.e., feed pressure variations of ± 2psi, fouled 5u prefilter, or shallow well pump performance.

A comparison between average measured flux rates from the matrices (February 28, March 29) and flux rates observed during the extended operation study, indicated that flux significantly declined for the extended study. Average flux rates for both studies are shown in Table 75.

Flux declines through the extended study may be attributed to membrane fouling due to organics, biological activity, or inorganic precipitation. However organic fouling is perhaps more likely due to the high organic content of the feed. The membrane matrices were conducted after 100 hours of operation at Olga, and the foulant layer may not have been fully established. Membrane foulant layers reduce permeate flux, and decrease solute rejection, due to greater concentration gradients at the surface. Permeate DOC at 105 psi-60% recovery was 1.6-mg/L during the relatively clean matrix study, compared to a value of 2.6-mg/L for the extended study. Other organic parameters, which are a function of DOC were correspondingly higher. Permeate values for inorganic parameters were also higher during the extended study.

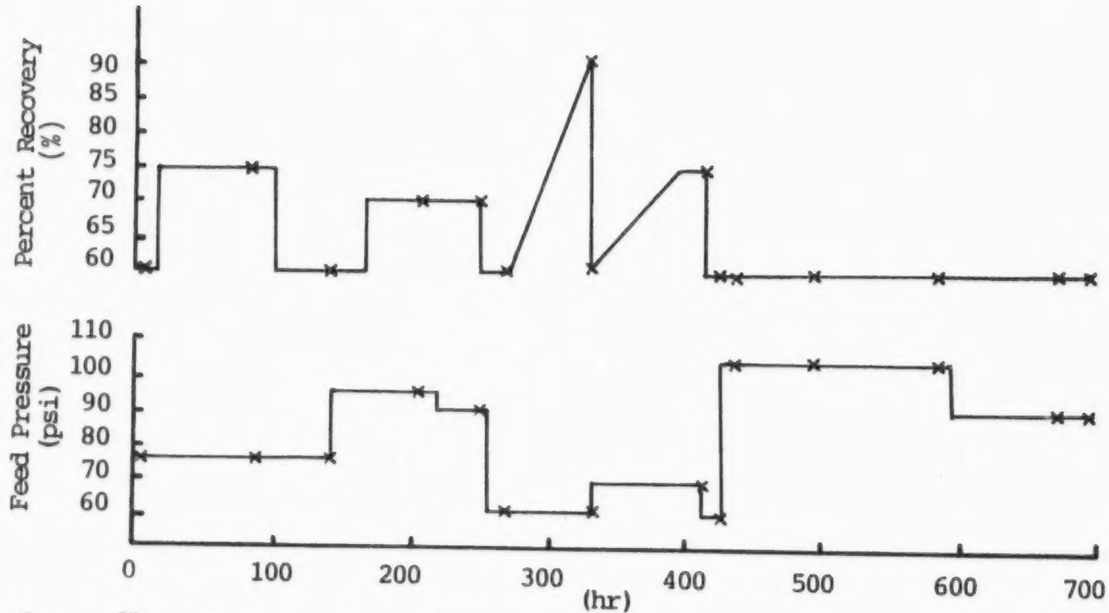


Figure 89 Operational settings of feed pressure and percent recovery versus hours of operation. Long term study at Olga, Florida.

Table 74. Flux Data for Olga N-50 Ultrafiltration Study.

Feed Pressure (psi)	No. of Points	Mean (gpd/ft²)	Std Deviation (gpd/ft²)	Range (gpd/ft²)
60	16	11.90	0.50	10.9 - 12.5
70	26	13.96	1.03	12.3 - 15.6
75	32	13.63	1.03	12.5 - 15.6
90	10	13.86	0.82	12.3 - 15.2
*95	26	18.46	2.05	16.0 - 22.3
105	35	15.02	1.50	12.5 - 17.9
120	6	17.1	0.32	16.9 - 17.6

*95 psi condition run 5.5 hours after cleaning.

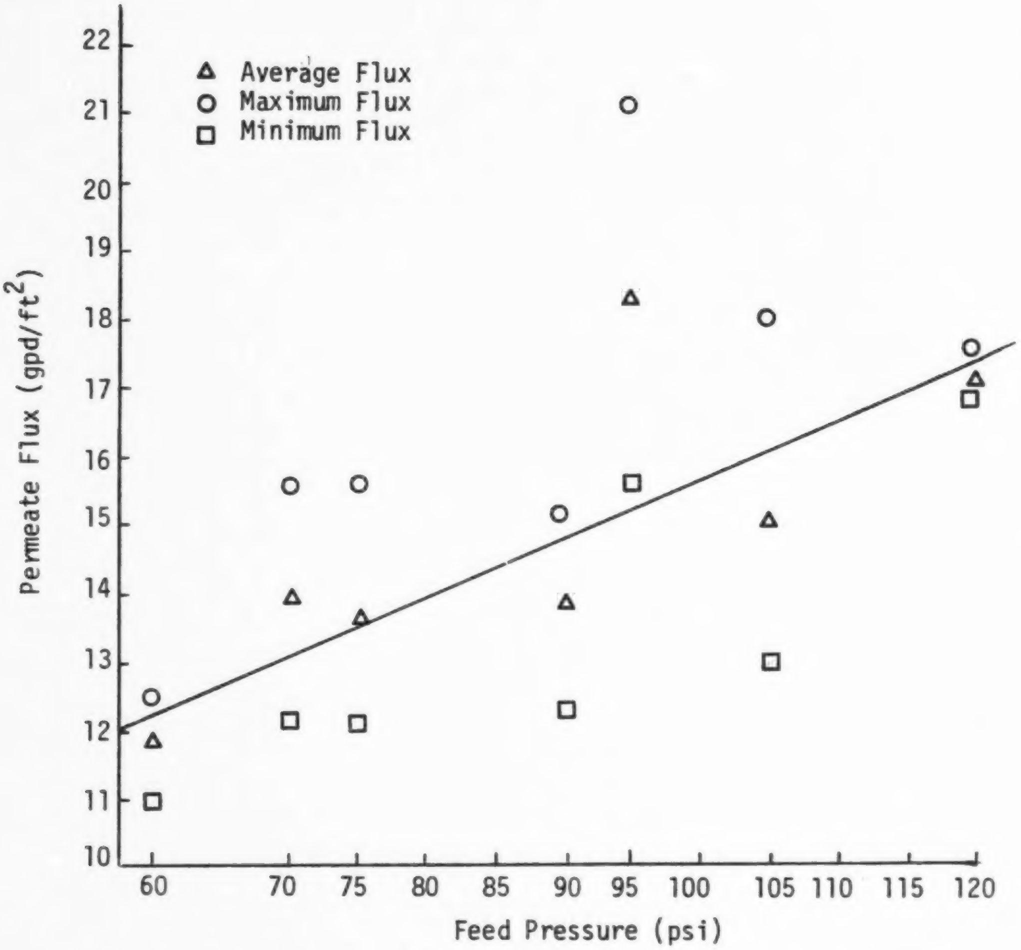


Figure 90 Permeate flux as a function of feed pressure, Olga N -50 Ultrafiltration Study

Table 75. Comparison between average permeate flux measured during matrix studies and extended operation, as a function of feed pressure Olga, Florida

Feed Pressure (psi)	Matrix Study Flux (gpd/ft²)	Extended Study Flux (gpd/ft²)
60	12.5	11.9
75	15.8	14.0
90	19.7	14.0
95*	N/A	18.5
105	21.8	15.0
120	24.2	17.0

*95 psi condition ititiated 5.5 hours after cleaning.

The comparison of matrix and extended study flux data yielded the following observations.

1) The permeate flux at 75 psi during the matrix study was 15.8 gpd/ft². The permeate flux at 105 psi during the extended study was 15.0 gpd/ft². An additional 30 psi of feed pressure was necessary to achieve the same flux under "semifouled" conditions.
2) The THM MCL was met at 75 psi-60% recovery after cleaning, and at 105 psi - 60% recovery during the extended study flux rates were 15.8 and 15 gpd/ft² respectively.
3) The 95 psi flux readings after cleaning averaged 18.5 gpd/ft². This is comparable to a flux reading at 85 psi during the N-50 matrix study. The matrix study was run under an optimal "clean" condition, yielding the best permeate quality attainable for the operating system.

Membrane surface conditions as well as applied pressure determined permeate flux. Under "fouled" conditions, fluxes were reduced over the range of operating pressures. The best quality permeate occured under "clean conditions" due to solute capture in reestablishing the foulant layer and at high flux, yielding low concentrations. Under partially fouled conditions, the best permeate quality was achieved at high pressures and low recovery ratios.

During the extended operation testing, the membrane unit was periodically flushed with "raw" prefiltered river water. Flushing took place by fully opening the reject value allowing increased raw flow to scour the reject side of the membrane surface. No chemicals were used during normal "flushing." Average flux increases after "flushing" were 0.5 gpd/ft². Flux increases were temporary, with fluxes declining to pre-flushing values within 1.0 hour.

The membrane was chemically cleaned once during the extended study (hour 151). Flux rates at 60 psi - 75% recovery had declined from 14.6 to 12.3 gpd/ft² during the first 150 hours of operation. The membrane was initially cleaned with a 0.2% solution of phosphoric acid, which was recirculated through the system for twenty minutes. The membrane was rinsed with 3.0 gallons of distilled water. A solution of 1% sodium hydroxide and 1% sodium EDTA (pH = 11.8) was then recirculated through the membrane for 30 minutes. Finally, the membrane was rinsed with 2.0 gallons of distilled water before being placed on-line.

Feed pressure was set at 75 psi with 60% recovery to duplicate the precleaning operating conditions. Permeate flux was 19.7 gpd/ft² compared to a pre-cleaning flux of 12.3 gpd/ft². Permeate flux declined to 18.3 gpd/ft² within 5.5 hours. Samples were collected prior to cleaning, 1.0 hour after cleaning, and 4.0 hours after cleaning, to determine changes in water quality. Organic parameters improved significantly, while inorganic parameters improved slightly. The N-50 membrane had been operated about 400 hours at other testing sites, previous to initial chemical cleaning.

After 5.5 hours of operation, operating conditions were changed to 95 psi at 60% recovery. Initial flux at this setting was 22.2 gpd/ft^2. Figure 91 depicts the decline in flux over 120 hours of operation. Permeate flux decline to 15.7 gpd/ft^2, a value near the expected average flux (see Figure 90).

"Clean" membranes are important to maintain maximum flux and water quality. Caloosahatchee River water is noted for high algae productivity and would require more elaborate pretreatment to reduce membrane fouling.

6.4.5.2 Organic Quality

Data for product water DOC, true color, 96 hr THM formation potential (dose = 5-mg/L Cl_2) and 96 hr TOX formation potentials are presented in Table 76. Figure 92(a), 92(b), 93(a), and 93(b) show percent removals of DOC, THM's, True Color, and TOXFP respectively, based on raw water values, as a function of feed pressure and percentage of recovery. Percent removals based on raw water values generally indicate that product water quality improves as pressure increases, and percentage of recovery decreases. Table A-14 presents calculated feed concentrations for DOC, color, and other inorganic parameters. Each organic parameter will be discussed below.

Dissolved Organic Carbon

The DOC data for raw, product and feed water, is plotted as for the corresponding feed pressure and percentage of recovery in Figure 94. The graph illustrates that percentage of recovery is the greatest factor determining product water DOC. The data point at 426 hours illustrates the effect of recovery on feed concentrations. The lowest product DOC values (≤ 2.0-mg/L) appear at the 75 psi - 60% recovery condition (hour 154 and 157). The membrane had been cleared at hour 153 due to a 16% decline in flux. Other low DOC values occured at 95 psi - 68% recovery (2.5-mg/L), 90 psi - 70% recovery (2.7-mg/L), and 60 psi - 61% recovery (2.0-mg/L). These samples were taken within 120 hours of operation after membrane cleaning. DOC values indicate the combined benefit of increased membrane flux and low recovery ratios. The effect of product recovery was also demonstrated by the 70 psi samples. DOC values for 90% recovery and 73% recovery were 12.8-mg/L and 6.1-mg/L respectively, even though raw DOC is greater at the 73% recovery condition. Differences in DOC values for the same operating conditions can be attributed to membrane fouling by raw water organics and inorganics (60 psi-60% recovery), and differences in raw DOC (105 psi-60% recovery).

Percent DOC removals based on an average raw value at 21.4-mg/L are plotted for the corresponding feed pressure and percentage of recovery in Figure 92a. At a constant raw DOC, removals would be greatest at high pressure and low recovery.

Percent DOC removal based on average raw DOC were 4 to 7% lower during the long term study compared to the test matrices, except at the 75 psi-60%

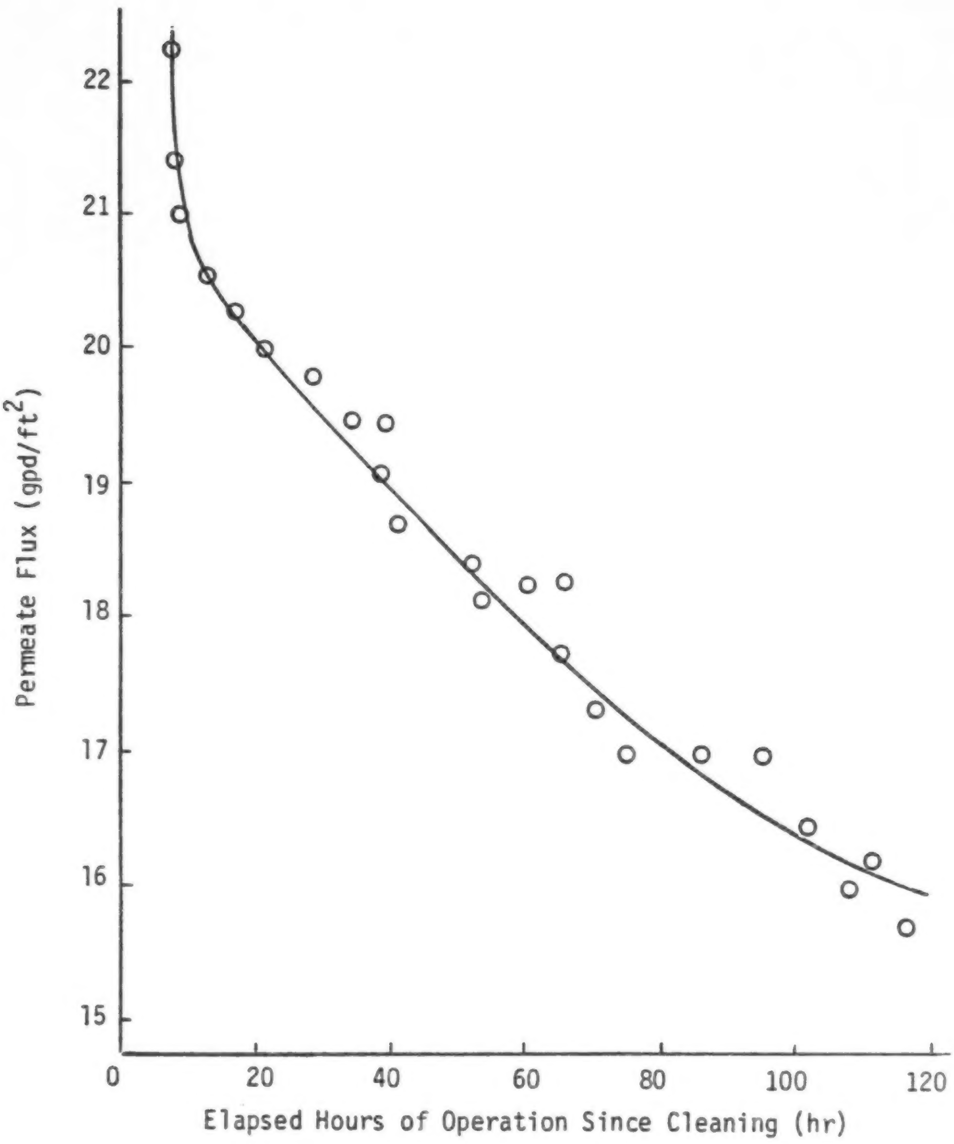


Figure 91 Permeate Flux at 95 psi feed pressure as a function of elapsed hours of operation since chemical cleaning. Olga N -50 Ultrafiltration Study.

Table 76. Olga N-50 Ultrafiltration Study Test Results (Organics)

Date (Hr. of Operation)	Sample Type/ % Removal	Feed Pressure (psig)	Percent Recovery (%)	DOC (mg/L as C)	True Color (cpu)	THM's 96 hr (μg/L)**	TOXFP 96 hr (μg/L as Cl)	Cl_2 Res 96 hr (mg/L)
4/11	Raw	75	60	18.0	58			
(5.5)	Product			1.8	2.5	158	169	0.8
	% Removal			90.2	95.7			
4/15	Raw	75	75	19.3	47			
(90.0)	Product			2.4	3.0	146	210	1.6
	% Removal			86.2	93.6			
4/18 *(1)	Raw	75	60	21.8	59			
(BC)	Product			1.7	2.0	134	168	2.1
(151)	% Removal			92.2	96.6			
4/18 (2)	Raw	75	60	21.8	59			
(AC) 1 hr.	Product			1.4	2.5	104	115	2.4
(154)	% Removal			93.6	95.7			
4/18 (3)	Raw	75	60	21.8	59			
(AC) 4 hr.	Product			1.4	1.5	72	130	1.8
(157)	% Removal			93.6	97.5			
4/23	Raw	95	68	18.1	47			
(218)	Product			2.5	5.0	173	273	1.2
	% Removal			86.2	89.0			
4/25 (1)	Raw	90	70	19.7	50			
(263)	Product			2.7	4.0	190	274	1.4
	% Removal			86.3	92.0			
4/25 (2)	Raw	60	61	19.7	47			
(271)	Product			2.0	3.0	197	220	.4
	% Removal			89.8	93.6			

*Membrane cleaned with solution of EDTA and NaOH at pH 11.0 per Filmtec procedure, BC - before cleaning. AC-after cleaning- ( ) numbers refer to data plot hours.

** Initial Cl_2 dose = 5 mg Cl_2/L.

(continued)

Table 76 (continued). Olga N-50 Membrane Extended Operation Test Results (Organics)

Date (Hr. of Operation)	Sample Type/ % Removal	Feed Pressure (psig)	Percent Recovery (%)	DOC (mg/L as C)	True Color (cpu)	THM's 96 hr (μg/L)**	TOXFP 96 hr (μg/L as Cl)	Cl_2 Res 96 hr (mg/L)
5/2	Raw	70	90	23.8	54			
(342)	Product			12.8	26	370	594	0
	% Removal			46.2	51.8			
5/6 (1)	Raw	63	60	24.7	54			
(426)	Product			4.9	16.0	241	280	1.1
	% Removal			80.2	85.7			
5/6 (2)	Raw	70	73	24.7	54			
(435)	Product			6.1	10	243	330	0
	% Removal			75.3	81.5			
5/7	Raw	105	60	24.6	54			
(442)	Product			3.7	4.0	218	237	+*
	% Removal			85.0	94.4			
5/9	Raw	105	60	20.9	59			
(498)	Product			2.6	3.5	105	152	+
	% Removal			87.6	94.1			
5/13	Raw	105	63	21.7	55			
(590)	Product			2.7	4.0	99	183	+
	% Removal			87.6	92.7			
5/16	Raw	90	60	22.1	62			
(680)	Product			2.0	4.0	141	177	+
	% Removal			91.0	93.5			
5/17	Raw	90	58	19.3	61			
(703)	Product			4.1	7.0	195	365	+
	% Removal			78.8	88.5			

* + Indicate chlorine residual present, but not determined by DPD method titration.

**Initial Cl_2 dose = 5 mg/L.

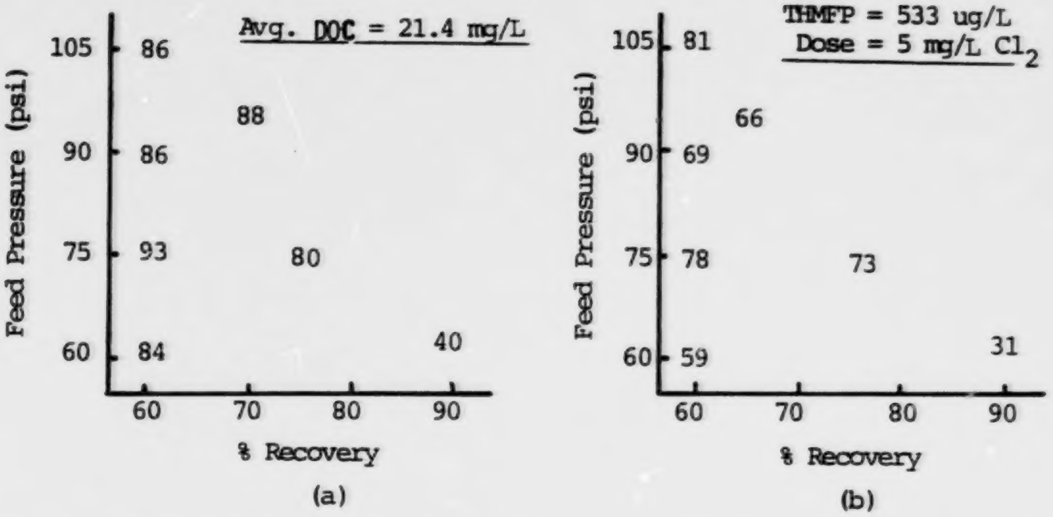


Figure 92 (a) percent DOC removal and (b) percent THM formation potential reduction, as a function of feed pressure and percent recovery - Olga N -50 Ultrafiltration Study.

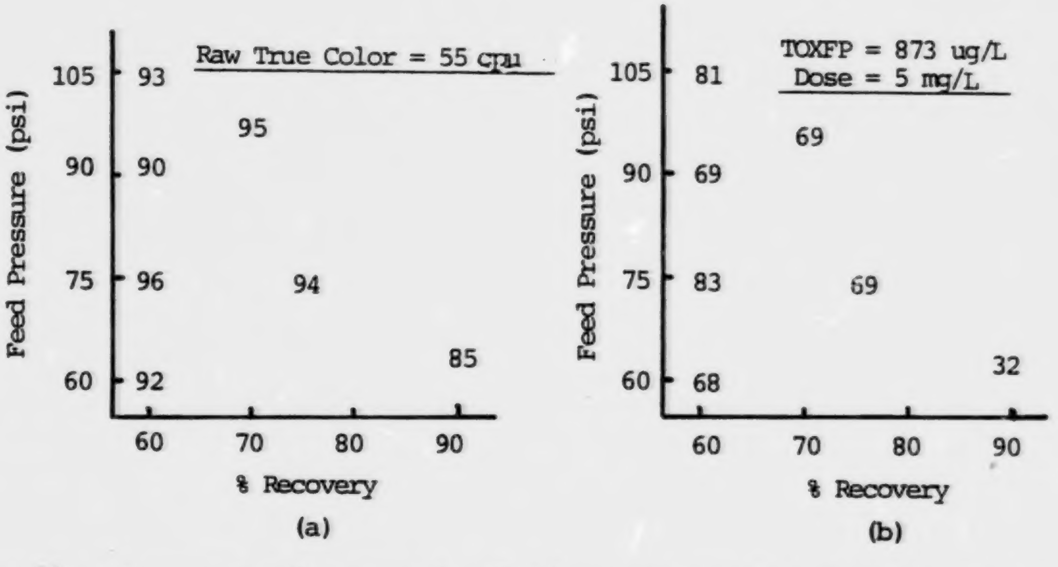


Figure 93 (a) percent color removal and (b) percent TOX formation potential reduction, as a function of feed pressure and percent recovery - Olga N -50 Ultrafiltration Study.

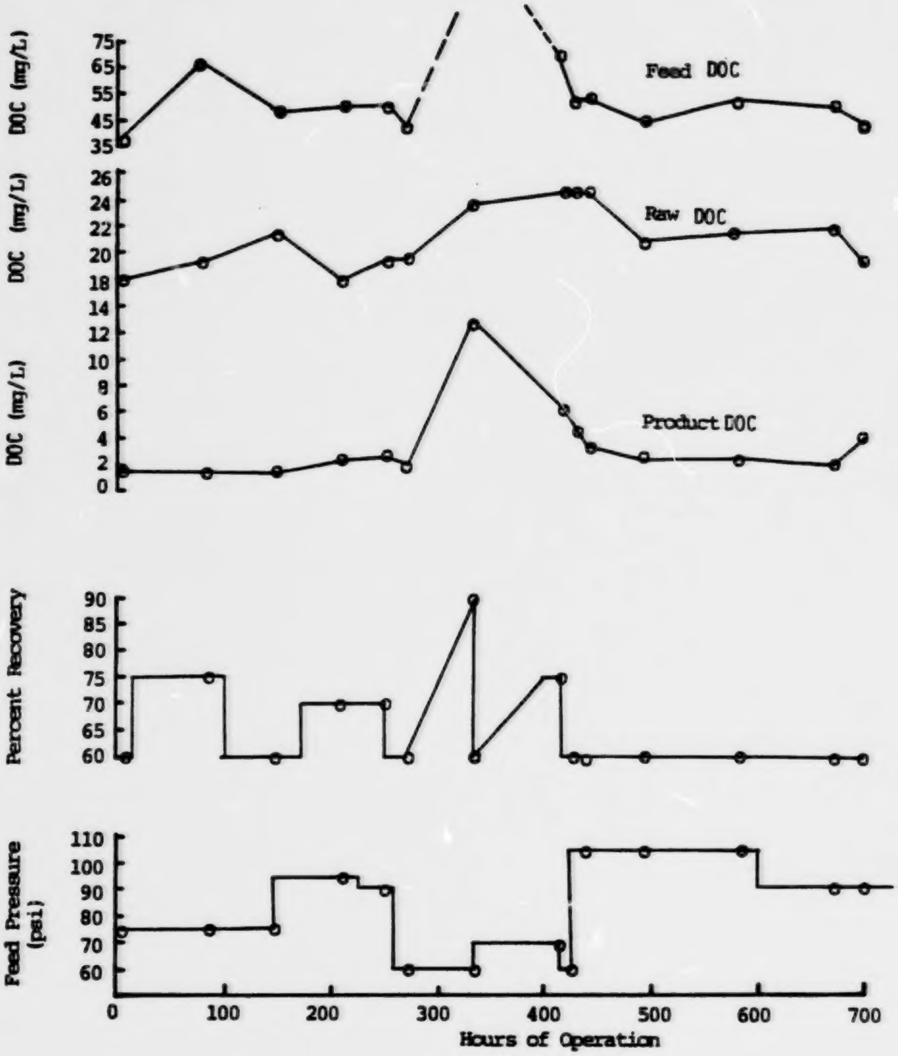


Figure 94 Raw, feed and product DOC as a function of feed pressure and percent recovery, during the Olga N -50 Ultrafiltration Study.

recovery "clean" operating condition. Percent DOC removals based on computed DOC feed concentrations ranged from 89.3% to 93.1%.

Trihalomethane formation potential is a function of DOC, chlorine dose, incubation temperature, pH and incubation time, as well as other variables. DOC and chlorine dosage were the most important factors in this study. THM formation potentials were developed using product water samples under the following conditions: chlorine dose = 5 mg Cl_2/L, incubation temperature = 25°C, incubation period = 96 hrs, and pH = 8.0. Ninety-six hour THMFP data is presented in Table 76. Chlorine residuals greater than 1.0-mg/L were present at lower product recoveries. Chlorine residuals greater than 2.0-mg/L were measured under the 75 psi - 60% conditions. Reducing chlorine dose and THM formation driving forces should result in lower values. The THM MCL of 0.10-mg/L was met at the 75 psi - 60% condition (clean membrane), and at the 105 psi - 60% condition near the end of the study (2 occasions).

Figure 95 shows 96 hr THM formation potential for the corresponding of raw water DOC and percentage of DOC removal. The lowest THM's occured at the upper right hand corner, indicating positive correlation between product DOC values and THM's. DOC concentrations (mg/L) rather than percent removals determine THM's.

Figure 92b shows percent THM's reduction as a function of feed pressure and percentage of recovery. Average THM's of the raw water (dose - 5-mg/Cl_2/L) was computed to be 533 µg/L as in the matrix study. The greatest reduction in formation potential occurs at high feed pressure and low percentage of recovery. Percent reduction in THM's was 5-15% lower during the long term study, compared to the matrix testing.

Data for 96 hour TOX formation potential is presented in Table 76. TOX data was correlated with DOC and THM's. The following relationship between product DOC and product THM's was calculated by linear regression:

$$\text{TOX } (\mu g/L) = 38.78 \text{ DOC } (mg/L) + 109.49.$$

The r^2 correlation coefficient is 0.866 (16 data points). THM's was correlated with TOXFP yielding the following equation;

$$\text{TOXFP } (\mu g/L)^2 = 1.47 \text{ THM's } (\mu g/L) - 13.87.$$

The r^2 correlation coefficient was 0.839 using 16 data points.

Percent reduction in TOXFP was calculated based on an average raw water TOXFP at 873 µg/L. The mean raw value was calculated using four values (1171, 840, 854, 630 µg/L). Percent reduction TOXFP is plotted in Figure 93b. The greatest reductions occur at high pressures and low recoveries. TOXFP reduction is also high at the 75 psi - 60% recovery condition, reflecting the clean state of the membrane at the time of sample collection.

Product water color values are presented in Table 76. Raw true color, product color, and feed color are plotted in Figure 96. The lowest product color values occur at 60% recovery at feed pressures ranging from 60 to 105

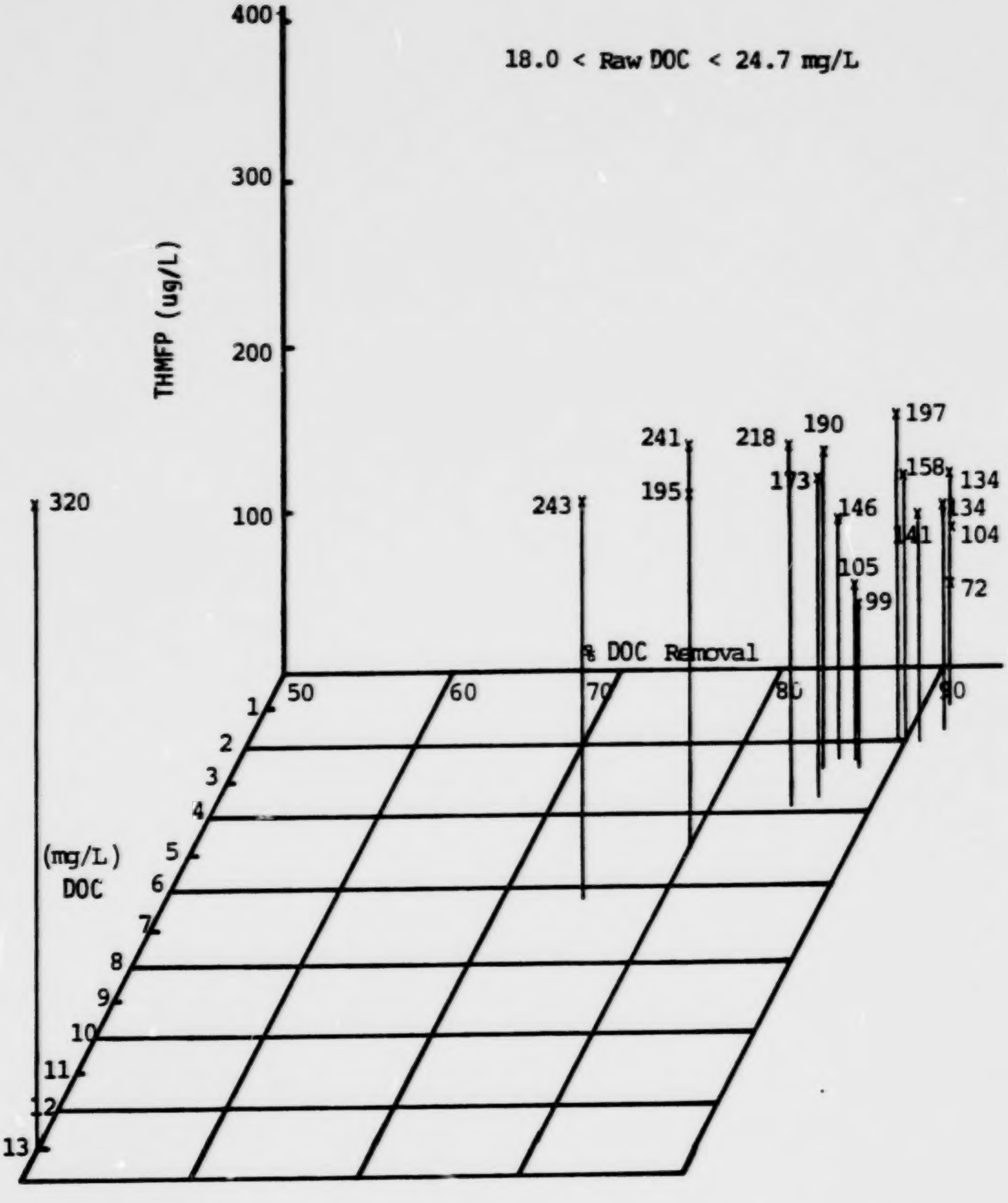


Figure 95 96 hr THMFP as a function of permeate DOC and percent DOC removal. Olga N -50 Ultrafiltration Study.

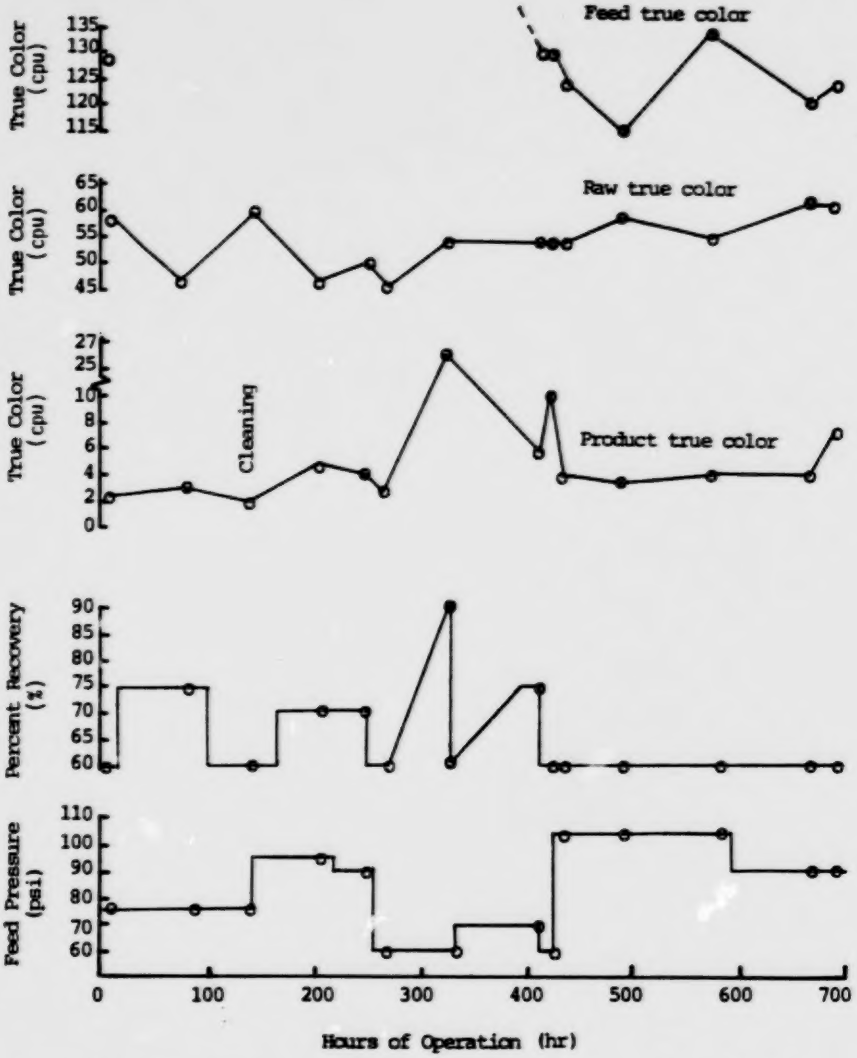


Figure 96 Raw, feed, and product true color values as a function of feed pressure and percent recovery during the Olga N -50 Ultrafiltration Study.

psi. Product color (cpu) was correlated with product water DOC. Linear regression analysis yielded the following relationship (r^2 = 0.963);

$$\text{true color (cpu)} = 2.039\ \text{DOC (mg/L)} - 1.483.$$

This indicates that product water true color is caused by dissolved organic compounds all of which do not produce color (negative intercept).

Product water true color values, were formatted as percent removals based on an average raw true color of 55 cpu. Percent color removal based on average raw color (Figure 93a), showed only minor variation at 50 to 75% recovery over all feed pressures. Color removals based on average raw true color were 2 to 5% lower in the long term study as compared to the matrixes, except for the 75 psi - 60% recovery condition, which was 1% higher.

In summary, product water THM's, TOXFP and true color are positively correlated to DOC levels. The major goal in membrane filtration is to reduce DOC levels. DOC was minimized under high pressure, low recovery operating conditions (105 psi, 60%), and under the "clean" condition (75 psi-60 psi).

6.4.5.3 Inorganic Quality

Inorganic parameters analyzed during the extended study were pH, alkalinity, total hardness, calcium hardness, total dissolved solids, chlorides, and sodium. All analyses except sodium were performed by UCF researchers. Sodium analysis was done by the Lee County environmental laboratory. Data for all inorganic parameters is reported in Table A-15. Figure 97(a), 97(b), 98(a), and 98(b) show percent removals of Total Hardness, total dissolved solids, chlorides, and alkalinity, respectively, as a function of feed pressure and percentage of recovery. Results for each parameter will be discussed separately.

The pH of the Caloosahatchee river source was fairly constant throughout the duration of the study. The pH range varied from 7.9 to 8.3 with the lower pH's occurring during the early part of the study. Product water pH ranged from 7.6 to 8.0 with the average reduction in pH of 0.36 pH units. Average product pH was 7.76. No trend, as a function of feed pressure or percentage of recovery, was noted. Decreasing pH is due to the rejection of HCO_3^- and non rejection of CO_2.

Alkalinity data is listed in Table A-15. Raw alkalinities varied from 176 to 149-mg/L as $CaCO_3$ during the test period. Percent reductions based on raw values were highest for the 75 psi - 60% recovery operating condition. At this setting, 65% reduction of alkalinity occurred, resulting in an average value of 62-mg/L as $CaCO_3$. Percent alkalinity reduction at the 105 psi - 60% recovery setting was 58.4 % with a mean value of 67-mg/L as $CaCO_3$. Rejection of alkalinity declined as percentage of recovery increased resulting in alkalinities of 106, 94, and 99-mg/L as $CaCO_3$ at 70 psi - 90% recovery, 75 psi - 75 percentage of recovery and 70 psi - 73% recovery, respectively. Figure 98b shows percent alkalinity reduction as a function of feed pressure and percentage of recovery. Alkalinity rejection

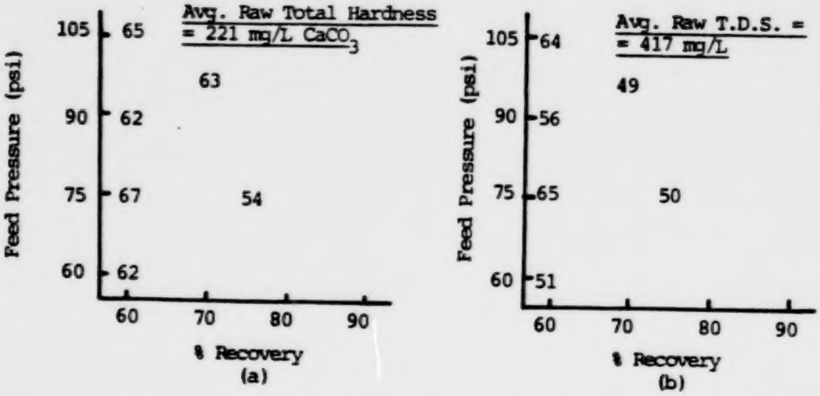


Figure 97 (a) Percent total hardness removal and (b) percent total dissolved solids removal as a function of Feed Pressure and percent recovery - Olga N -50 Ultra Filtration Study.

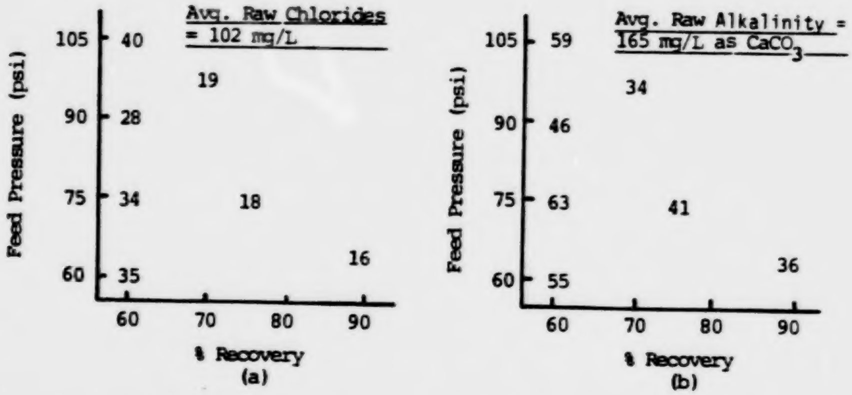


Figure 98 (a) Percent chloride removal and (b) percent alkalinity reduction as a function of feed pressure and percent recovery - Olga N -50 Ultrafiltration Study.

declines with increasing percentage of recovery, based on an average alkalinity of 165-mg/L as $CaCO_3$. Alkalinity rejection was 10 to 15% lower during the extended study compared with the test matrices.

Total hardness data is shown in Tables A-14 and A-15. Raw total hardness varied from 245 to 206-mg/L as $CaCO_3$ during the test period. Product, raw and computed feed total hardness are shown in Figure 99. The lowest product total hardness values occured at the 60% recovery ratio and were relatively insensitive to feed pressure. Average total hardness values at 60% recovery were 85, 73, 84, and 77-mg/L as $CaCO_3$, at 60, 75, 90 and 105 psi, respectively. Average percent TH removals at 60% recovery were 64% based on raw water concentration. Total hardness concentrations at 70 psi declined as percentage of recovery increased, with values of 102-mg/L at 75% recovery and 163-mg/L at 90% recovery. Percent total hardness removal, based on an average total hardness of 165-mg/L as $CaCO_3$, is plotted in Figure 97a. Rejection declines as percentage of recovery increases. Percent rejection based on calculated feed concentrations averaged 80% with a range of 73.9 to 86.4%. Percent removals were 10 to 15% lower during the extended study, compared to the test matrices.

Table A-15 presents calcium hardness data. Calcium hardness averages 86% of permeate total hardness. The ratio of calcium hardness to total hardness increased as percentage of recovery increased, averaging 74, 88, and 96% at recoveries of 60,70, and 90% respectively. Rejection of calcium hardness followed the same pattern as total hardness, with permeate concentrations increasing with increasing recovery.

Product and raw TDS values are presented in Table A-15. Raw, product and computed feed TDS values are shown in Figure 100. Lowest TDS values occured at 60% recovery with the exception of the sample at 271 hours (60 psi - 61% recovery). Product water TDS increased with increasing recovery at constant pressure, with average values at 75 psi of 145-mg/L at 60% recovery, and 207-mg/L at 75% recovery. Percent TDS removal based on computed feed concentrations averaged 78% with a range of 83.6% to 71.7%. Figure 97b shows percent TDS removal as a function of feed pressure and percentage of recovery. Average raw TDS was 417-mg/L. Percent TDS removal decreased with increasing recovery, and was relatively insensitive to feed pressure.

Raw and product chloride data is presented in Table A-15. Product, raw, and computed feed values are plotted in Figure 101 for the corresponding feed pressure and percentage of recovery. Raw water chloride values averaged 102-mg/L with a range of 85 to 122-mg/L. Product water chloride values for 60% recovery averaged 68, 74 and 62-mg/L at feed pressures of 75, 90, and 105 psi, respectively. Thus, chloride rejection was relatively insensitive to feed pressure. Percent removals based on raw values averaged 31.4% at 60% recovery (67-mg/L), 25% at 70% recovery, (87-mg/L) and 16% at 90% recovery (86-mg/L). Figure 98a indicates percent removal based on an average raw value of 102-mg/L. Percent removal decreases as percentage of recovery increases. Chloride rejection based on computed feed concentrations averaged 50% with a range of 41.5 to 61%. Chloride rejection was 10 to 17% lower than the matrix studies.

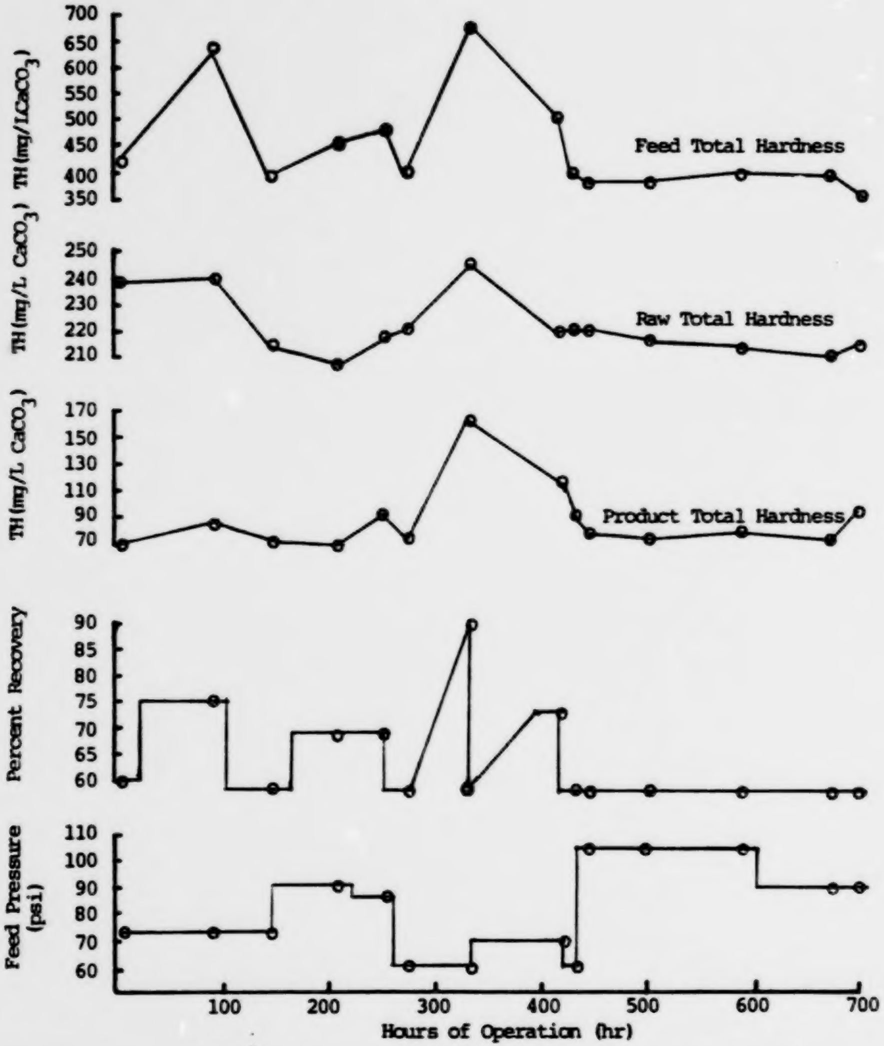


Figure 99 Raw, feed, and product total hardness values as a function of feed pressure and percent recovery, during the Olga N -50 Ultrafiltration Study.

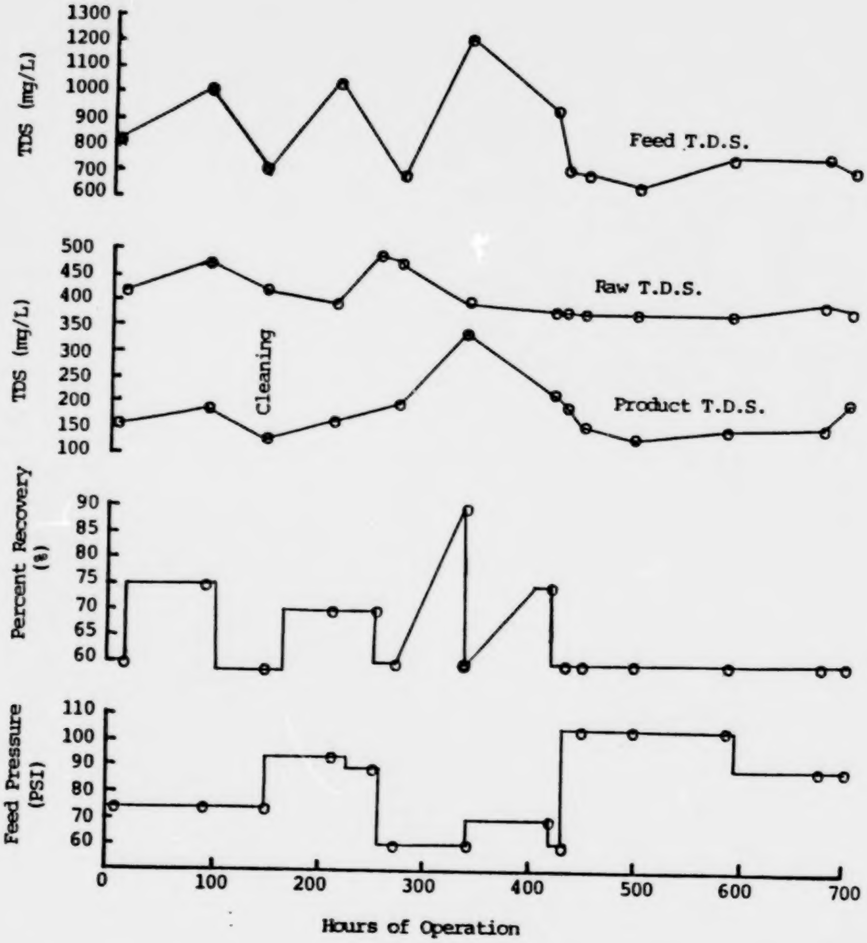


Figure 100 Raw, feed and product total dissolved solids for the corresponding feed pressure and percent recovery, during the Olga N -50 Ultrafiltration Study.

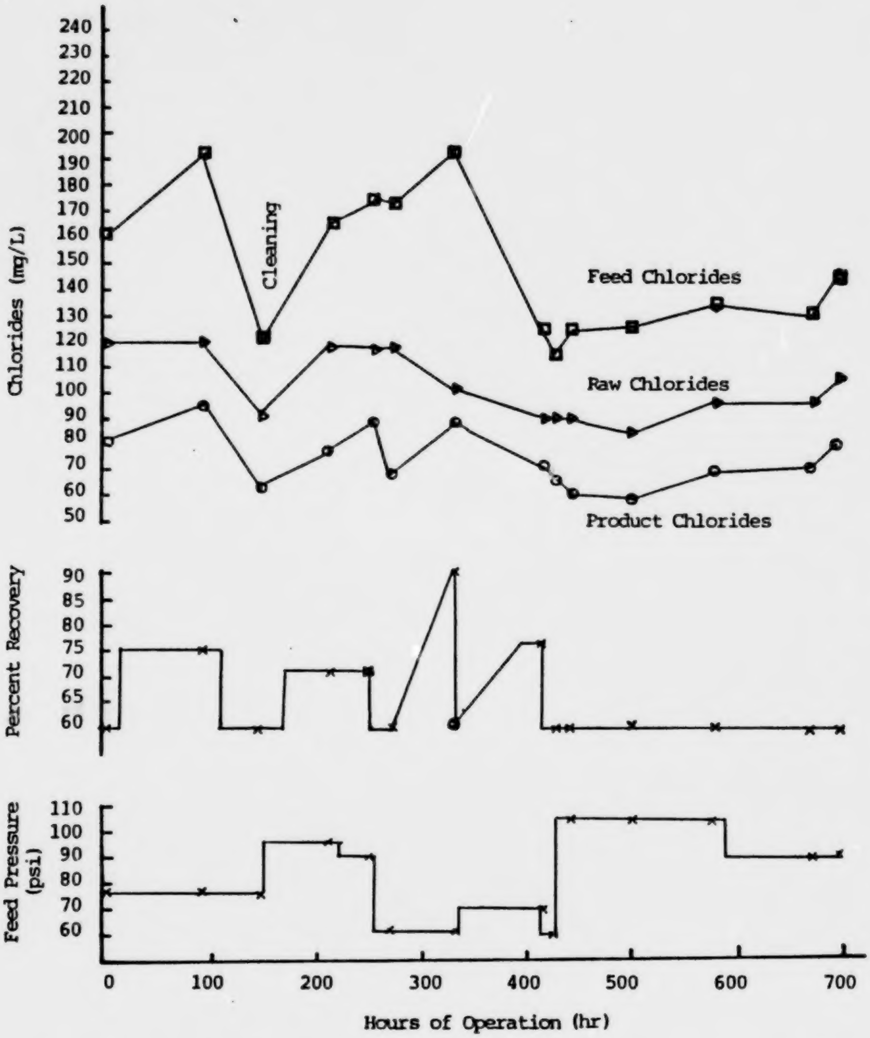


Figure 101 Raw feed and product chloride values as a for the corresponding pressure and percent recovery during the Olga N -50 Ultrafiltration Study.

Table A-15 lists raw and product water data for sodium. Raw water concentrations varied from 48 to 74-mg/L. Percent removals based on raw concentrations were highest at 60% recovery with removals of 26%, 25%, 35%, and 17% at feed pressure of 105, 90, 75, and 60 psi respectively. Thirty five percent removal at 75 psi - 60% recovery represents the "clean" membrane condition. Percent removals at 70 psi - 73% recovery and 70 psi - 90% recovery were 8.3% and 16.1% respectively, based on single data points.

In summary, rejection of inorganic compounds is largely dependent on percentage of recovery and only slightly dependent on feed pressure. The ranges for product pH (7.6 to 8.0), alkalinity (50 - 99-mg/L as $CaCO_3$), and calcium hardness (54-110-mg/L as $CaCO_3$) at recoveries up to 75%, are desirable for potable water. Chloride rejection based on raw values, averaged 29% for all operating conditions, but was variable (16-44%). Sodium rejection based on raw values, averaged 24% ranging from 38% to 8%. The importance of a "clean membrane" is illustrated by consistently high inorganic rejection at the 75 psi - 60% recovery condition.

6.4.6 Summary

The membrane selection study using seven membranes provided by three manufacturers, yielded two membranes, the Filmtec N-50 (MW = 400) and the Filmtec BW 30 R.O. membrane (MW = 100), which produced a high quality permeate, low in DOC (1.2-mg/L), with color less than 3.0 cpu. The N-50 was selected for extended study, because it produced a stable water (pHs = 7.7 at pH = 8.1) at a 65 psi feed pressure. In comparison, the BW 3030 produced a highly corrosive permeate at 190 psi.

Size fractionization using seven membranes with MW cutoffs of 100 to 40,000 MW yielded the following information about organic and inorganic quality of the Caloosahatchee River raw water source. Ninety-four percent (by mass) of the DOC had MW greater than 500 with 70 percent between 2000 and 20,000 MW. Ninety-six percent of the true color component had MW greater than 500. Eighty percent at the total hardness component had MW greater than 500. Seventy-six percent of total alkalinity had MW greater than about 500 MW, with 54% between 500 and 10,000 MW. Sixty-three percent of the chlorides had MW greater than 500 with 57% between MW's of 500 and 10,000.

A N-50 pressure (60-120 psi) - percentage of recovery (50 to 80%) matrix was conducted on February 28, 1985 using a "clean" N-50 membrane. Generally permeate quality was "best" at high feed pressures (105-120 psi) and low percentage of recovery (50-60%). The lowest permeate DOC (1.0-mg/L) occurred at the 120 psi - 50 and 60% settings. DOC values for 105 psi at 50, 60, and 70% recoveries averaged 1.6-mg/L, from a raw DOC of 19.6-mg/L. Ninety-six hr THM's (Cl_2 dose = 5-mg/L) for the stated settings, met the THM MCL with chlorine residuals greater than 2.0-mg/L. Permeate color values ranged from 3.5 cpu at 60 psi to 1.5 cpu at 120 psi, and were relatively insensitive to percentage of recovery.

The matrix inorganic parameters (pH, alkalinity, total hardness, calcium hardness, chlorides, and sodium) were lowest at high pressures

(105-120 psi) and low recoveries (50-60%). Average permeate values for all operating conditions were pH = 7.7, a total alkalinity of 44-mg/L $CaCO_3$, a total hardness of 56-mg/L $CaCO_3$, a calcium hardness of 42-mg/L $CaCO_3$, TDS of 172-mg/L, chlorides of 67-mg/L, and sodium of 42-mg/L. Matrix permeate flux ranged from 12.5 to 24.2 gpd/ft² over a 60 to 120 psi pressure range. These "high" fluxes reflect a relatively clean membrane used in the matrix. A second N-50 pressure (75-105 psi) - percentage of recovery (60-70%) matrix was conducted under the same flux conditions. Permeate DOC averaged 1.3-mg/L at 60% recovery and 1.6-mg/L at 70% recovery over the range of pressures. Color values were slightly lower than the previous matrix. Inorganic parameters were approximately the same as the first matrix.

The extended operation study on the N-50 quantified the effect of membrane surface condition on permeate quality and flux. The flux rates were lower at all pressure settings, compared to the two matrices. Fluxes at 60 psi and 105 psi averaged 11.9 and 15 gpd/ft² respectively. Due to membrane fouling, flux at a particular setting always declined over time. The "best" permeate DOC occurred at 75 psi - 60% recovery under clean conditions (4 hrs after cleaning) and at 90 and 105 psi at 60% recovery, under partially fouled conditions. DOC values were 1.4, 2.0, and 2.6-mg/L respectively. The lowest percentage of recovery (60%) used in the study yielded the lowest DOC values. Fluxes at the three "best" conditions averaged 15.5 gpd/ft², a relatively high flux during the extended study. DOC rejections averaged 4 to 7% lower than the matrices. DOC removals based on computed feed concentrations were greater than 89%.

THM MCL of 0.1-mg/L was met at the 75 psi - 60% recovery "clean" setting and at the 105 psi-60% recovery condition. Ninety-six hr THM's's at 5 mg Cl_2/L were 72 and 99 µg/L respectively, with Cl_2 residuals greater than 1.0-mg/L. Failure to meet the MCL at higher recoveries was due to higher DOC's averaging 2.6, 4.2, and 12.8 at 70, 75 and 90% recoveries, respectively, over the range of feed pressures. The TOXFP averaged about 1.5 times THM's and was positively correlated with permeate DOC (r^2-0.87). The largest reductions in TOXFP occurred at high pressures and low percent recoveries.

Permeate pH during the extended operation decreased an average of 0.36 pH units from the raw water value. No trend as a function of feed pressure or percentage of recovery was noted. Average permeate pH was 7.8. Alkalinity rejections were highest at the "clean" condition (75 psi-60%) and at the 105 psi-60% setting. Average alkalinity at these settings was 65-mg/L $CaCO_3$, a reduction of 61% from the average raw value. Rejection of alkalinity declined as percentage of recovery increased for a constant feed pressure. Alkalinity rejection was 10-15% lower during the extended study, compared to the test matrices.

Permeate total hardness was reduced to an average value of 78-mg/L $CaCO_3$ at 60% recovery for feed pressures of 60, 75, 90, and 105 psi. Thus, total hardness rejection was relatively insensitive to feed pressure. Percent removals were 10 to 15% lower during the extended study, compared to the matrix results. Calcium rejection followed the same trends as TH, averaging 86% of the permeate TH value. The ratio of CaH to TH increased as percentage of recovery increased, at a constant feed pressure. Permeate TDS was in the 150 to 225-mg/L range with the lowest values occurring at low recoveries. TDS was relatively insensitive to feed pressure. Percent removals based on computed "feed" concentration averaged 78%.

The N-50 UF membrane can produce an acceptable potable water low in DOC, having desirable hardness and alkalinity. Test matrices indicate that low MW UF (MW < 400) membranes can produce water meeting the THM MCL over the short run. Permeate quality over the long run is a strong function of membrane surface condition, as well as feed pressure and percentage of recovery. Extensive pretreatment of the Caloosahatchee River would be necessary to maintain permeate quality over reasonable operating periods. Membrane performance could be maintained, if routine cleaning procdures were adopted.

6.4.7 VOG Membrane Tests

6.4.7.1 Optimum Membrane Testing

As earlier stated three "optimum" membranes were tested on a bench-scale level at the VOG site. The data for the BW30, N-50 and U90-G10 membranes are given in Tables 77, 78, 79, and 80. Table 77 gives operating conditions for these membranes. Table 78 provides the water quality data for the basic organic and inorganic parameters while Tables 79 and 80 provide the THM formation curve and chlorine residual curve data. It should be remembered that the BW 30 is a RO membrane with a 100 MW cutoff pore size while the N-50 and the U90-G10 are UF membranes with molecular cutoff pore sizes of 400 and 2000, respectively.

As can be seen from the data, the BW30 produces a permeate water that has the characteristics of a RO water. The product water DOC value of 0.28-mg/L represents a 98% reduction from the raw water value, while color has been removed to a level below the detection limit. The low organic levels are further illustrated in the THM's and chlorine residual data in Tables 79 and 80. It was found that a 5-mg/L dose would leave a suitable residual at the end of 192 hours and produce THM's values well below the current MCL. Inorganic parameters are also removed to relatively low levels with reduction from the raw water values of 92% and 88% for total hardness and alkalinity, respectively. It should be noted that the product water levels of alkalinity (28-mg/L as $CaCO_3$) and total hardness (20-mg/L as $CaCO_3$) would produce a very "soft" water with virtually no buffering capacity. Such a product water alone would not be desirable for potable purposes on a long term scale without the addition of alkalinity and hardness.

Further inspection of the data shows that the N-50 membrane also achieved excellent organic removals with product water DOC values of approximately 1-mg/L as C and color values of 1 cpu. These relatively low values can also be seen in the THM's and chlorine residual values. With a chlorine dose of 10-mg/l giving a substantial residual after 192 hours, the THM's value is well below the current MCL of 0.10-mg/L. Also from a potable water view, the alkalinity and total hardness values are present in

TABLE 77. Optimum Membrane Testing Conditions

Membrane	Feed Pressure (psi)	Product Recovery Rate (%)
BW30	190	60
N50	60	65
V90-G10	100	80

Table 78. VOG Optimum Membrane Testing Data and Product Water Parameter Removals

Sample and Membrane	pH	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L $CaCO_3$)	CaH (mg/L $CaCO_3$)	TH (mg/L $CaCO_3$)	Turbidity (NTU)
VOG U90-G10	7.1	6.34	21	207	207	223	0.33
Raw	7.2	14.70	60	226	225	234	0.33
% Removal	--	57	65	8	8	5	0
VOG BW30	8.2	0.28	0	28	12	20	0.11
Raw	7.1	13.65	61	226	220	236	0.31
% Removal	--	98	99	88	95	92	65
VOG FT-50	6.8	1.03	1	88	73	78	0.18
Raw	7.10	15.60	59	222	220	235	0.27
% Removal	--	93	98	60	67	67	33

Table 79. VOG Optimum Membrane Testing - THMFP Formation Curve (all values in μg/L)

Site	Membrane	Initial Cl_2 Dose (mg/L Cl_2)	Time (hrs.) 1	3	24	48	72	96	192
VOG	BW30	5	14	13	11	11	28	21	18
VOG	BW30	10	11	11	11	11	32	--	28
VOG	FT-50	5	4	16	11	16	35	31	31
VOG	FT-50	10	11	16	16	18	39	--	23
VOG	U90-G10	5	48	61	99	84	91	90	--
VOG	U90-G10	10	85	133	221	225	326	451	430

Table 80. VOG Optimum Membrane Testing - Chlorine Residual Curve Data (all values in mg/L-Cl_2)).

Site	Membrane	Initial Cl_2 Dose (mg/L Cl_2)	Time (hrs.) 1	3	24	48	72	96	192
VOG	BW30	5	3.4	3.4	3.3	2.2	1.6	1.4	0.3
VOG	BW30	10	7.4	7.4	5.8	6.2	5.4	3.4	2.2
VOG	FT-50	5	2.7	2.0	1.6	1.4	0.9	0.8	0
VOG	FT-50	10	NA	8.5	8.0	7.1	6.5	3.4	1.7
VOG	U90-G10	5	0.36	0	--	--	--	--	--
VOG	U90-G10	10	5.10	5.0	3.4	3.1	2.4	1.3	0.34

levels (88 and 78-mg/l as $CaCO_3$, respectively) that are considered desirable for potable waters.

The water quality for the U90-G10 membrane exhibits relatively much higher organic and inorganic levels. Approximately one-half of the raw water DOC is still present and the color value of 21 cpu would require chlorine bleaching. The relatively high organic values are confirmed in the THM's value of 430 μg/L for the 192 hour sample at an initially required chlorine dose of 10-mg/L. Finally, it should be noted that the total hardness value of 223-mg/l as $CaCO_3$ would be considered a "hard" water, though potable.

Given the stated molecular cutoff weights the results found are not unexpected. It would appear that the organic parameters are present in molecular weights greater than those found for the inorganic parameters. For each membrane the organic parameter removal is greater than the inorganic parameters on a percentage basis. Again this would be expected given the nature of fulvic and humic acids with their large molecular weights.

From the data it was apparent that the BW 30 and the N-50 membranes were capable producing a product water that would meet the THM MCL. Thus the decision between the two membranes was made on other considerations.

The N-50 membrane was favored at the VOG site because of cost and quality. The required feed pressure for the N-50 membrane was approximately 150 psi less than the BW 30. For a 0.5 MGD facility at VOG this would be a difference of approximately 26,500 kw-hrs/year. Also construction materials demands for a high pressure RO operation would be more severe. For potable purposes, the water produced by the N-50 membrane exhibits an alkalinity and total hardness that is desirable. The alkalinity available would buffer against pH changes induced by chlorination. The hardness present in the water would contribute to desirable taste characteristics. The water produced by the BW30 membrane would be subject to severe pH changes due to chlorination and would result in a corrosive water.

6.4.7.2 Pilot Plant Studies

The results of the February 2, 1985 test matrix are given in Tables 81 82, 83, 84, and 85. In this matrix feed pressure varied from 80 psi to 110 psi while product recovery rates varied from 50% (no recirculation) to 90%. Examination of the data in Table 81 shows that numerically small ranges are found for color and THM's with larger ranges found for TDS and TH. The color values have a range of only 1.0 cpu with THM's having a range of 8.0 μg/L - 28 μg/L. The range for TDS is 136-mg/L - 256-mg/L while TH ranges from 98-mg/L as $CaCO_3$ to 195-mg/L as $CaCO_3$.

Examination of Tables 82, 83, 84, and 85 show more clearly the effect of feed pressure and product recovery rate. In Table 82 color does not exhibit any definite patterns. This is probably due to the relative imprecise measurements at these low levels. The trends in THM's shown in Table 83 are also rather tenuous with slight increases seen for each feed pressure as product recovery rate increases. Trends for low values as

Table 81. VOG Test Matrix - Product and Raw Water Quality
(Testing on 2-2-85)

Feed Pressure (psi)	Product Recovery Rate (%)	Total Dissolved Solids (mg/L)	Total Hardness (mg/L as $CaCO_3$)	True Color (cpu)	Trihalomethane Formation Potential (μg/L*)
80	50	146	109	0.5	10
80	66	198	128	0.5	14
80	75	205	135	0.5	17
80	90	259	195	1.5	28
90	50	137	98	0.5	9
90	66	171	135	0.5	15
90	75	180	135	0.5	12
90	90	193	150	0.5	17
100	50	136	101	0.5	11
100	66	160	124	0.5	13
100	75	169	139	1.0	15
100	90	188	135	0.5	11
110	50	147	101	0.5	8
110	66	214	120	1.0	14
110	75	200	130	0.5	11
110	90	219	139	0.5	11
Raw #1	--	336	244	58	NA
Raw #2	--	367	237	58	NA
Raw #3	--	332	237	57	NA

*THM's from samples dosed at 7 mg/L Cl_2 and incubated for 48 hours at 25°C and pH 8.0.

Table 82. VOG Test Matrix True Color vs. Feed Pressure and Product Recovery Rate
Raw Water Average = 58 cpu

Feed Pressure (psi)	Product Recovery Rate %			
	50	66	75	90
80	0.5	0.5	0.5	1.5
90	0.5	0.5	0.5	0.5
100	0.5	0.5	1.0	0.5
110	0.5	1.0	0.5	0.5

Table 83. VOG Test Matrix Trihalomethane Formation Potential vs. Feed Pressure and Product Recovery Rate
Raw Water Average (without chlorination) = 0 μg/L

Feed Pressure (psi)	Product Recovery Rate %			
	50	66	75	90
80	10	14	17	28
90	9	15	12	17
100	11	13	15	11
110	8	14	11	11

Table 84. VOG Test Matrix Total Dissolved Solids vs. Feed Pressure and Product Recovery Rate
Raw Water Average = 345 mg/L

Feed Pressure (psi)	Product Recovery Rate %			
	50	66	75	90
80	146	198	205	259
90	137	171	180	193
100	137	160	169	188
110	147	214	200	219

Table 85. VOG Test Matrix Total Hardness vs. Feed Pressure and Product Recovery Rate
Raw Water Average = 239 mg/L as $CaCO_3$

Feed Pressure (psi)	Product Recovery Rate %			
	50	66	75	90
80	109	128	135	195
90	98	135	135	159
100	101	124	139	135
110	101	120	130	139

experienced here are more difficult to observe because of the nature of THM measurement.

More definitive trends are seen in the TDS values (Table 84). A definite increase is seen as product recovery rate increases within each feed pressure level. Also decreases are generally seen in each product recovery rate level with increasing feed pressure with the exception of the 110 psi feed pressures. At this feed pressure a definite and usually substantial increase in TDS is experienced. Similar results are exhibited for TH (Table 100) as product recovery rate increases. However, TH does not vary substantially or consistently with increasing feed pressure except for the 90% product recovery rate level.

The trends noted in the inorganic parameters are probably due to a combination of concentration polarization and membrane behavior factors. At the stated nominal molecular weight cutoff (400 MW) the membrane is in a region shared by the UF and RO processes (Weber 1972). It is reasonable to expect the membrane to exhibit separation properties that are germane to each process. Descriptions of the concentration polarization phenomenon has been given by a number of authors (Badenhop et al., 1970, Michaels et al., 1971, DiFilypi and Goldsmith, 1970). In this process solute is brought to the membrane surface by convective forces. This results in a solute concentration increase at the membrane surface that increases in thickness and concentration until the back transport rate equals the convective transport. The end result is a surface layer with a solute concentration relatively greater than the bulk solution. Such a phenomenon has several effects on both solute and solvent flux:

1) The solvent flux will be decreased from the pure solvent flux rate. This will occur from two effects whose relative importance is based on the degree of character the membrane exhibits for RO or UF operation. In RO the increased solute concentration would increase the effective osmotic pressure and thus decrease the solvent flux (from Weber 1972).

   $$Fw = Wp\ (\Delta P - \Delta\pi)$$

   with Fw = flux of water

   Wp = coefficient of water permeation

   ΔP = transmembranic pressure difference

   $\Delta\pi$ = difference in osmotic pressure.

   For UF the decrease is due to the effective formation of an additional hydraulic resistance by the concentrated solute layer (Michaels et al., 1971). Such a retardination of flow can be modeled as two hydraulic resistances in series (solute layer and membrane)

2) The solute flux is increased by the concentrated solute layer. (from Weber 1972)

$Fi = kp \Delta Ci$

with Fi = flux of solute i

kp = coefficient of permeability for solute i

ΔCi = transmembranic solute concentration differences.

Thus, as the effective concentration on the upstream side of the membrane increases, the difference increases as does flux of the solute. For UF the predominant solute transport mechanism is molecular sieving and the effects of the concentration polarization could be mixed. The concentration effect would bring a great number of molecules to the membrane surface thus increasing the change of passage. However the concentrated nature of the layer could also impede solute movement or allow a greater chance of chemical interaction between solute species. Complexition, chelation or precipitation are three possible reactions under the elevated concentration conditions. Thus solute behavior in the UF mode would be difficult to predict.

3) While the degree of solvent passage retardation or solute passage augmentation might be difficult to predict, it would be reasonable to assume that the concentration polarization will increase product water solute concentration. This will be due to a decrease in solvent (water) and a concurrent increase in solute (i.e., total dissolved solids).

Having described concentration polarization and its causes, the relation of this phenomenon to the feed pressure and product recovery rate should be explored. An increase in product recovery rate translates into an increased use of concentrate (reject) water as recirculation in the feed stream. This increased proportion of relatively higher solute concentration water serves to increase the bulk solution solute concentration in the feed stream. It can be reasonably expected that an increased bulk stream solute concentration will increase the concentration polarization layer effects (aka "gel" layer) on solute concentration. Furthermore it then follows that as the product recovery rate increases so will the product water solute concentration (with all other factors constant).

The effect of an increase in feed pressure will affect both solute and solvent flow. An increase in feed pressure will increase solvent (water) flux by increasing transmembranic pressure difference. This would have the effect of "diluting" the solute concentration on the product side. However an increased feed pressure will increase the convective movement towards the membrane and result in an increased concentration polarization layer. The relative effects of the opposing forces would be expected to be a function of the particular feed stream.

With the above discussion, which will be germane to a significant proportion of the membrane discussion, the matrices for TDS and TH may be explained. As can be seen in Table 84 product water TDS increases with increasing product recovery rates. This is consistent with the discussion of concentration polarization. The effect of increasing feed pressure is interesting in that for the range of 80 psi to 100 psi, the increased feed pressure serves to increase solvent flux more than increase solute flux. However, at 110 psi feed pressure it would appear that the solute passage effect is dominating the solvent passage effect.

Examination of Table 85 shows a similar behavior for TH for the product recovery rate effect. However the effect of increased feed pressures is less defined with general but not consistent declines in product water TH concentration for the 50%, 66% and 75% product recovery rates. The 90% product recovery rate level shows a similar trend to the TDS values. Such patterns do not have an readily apparent cause, a possible explanation could involve unknown relative solvent - solute interaction factors.

In discussing the apparent lack of patterns within the THM's and true color values (see Tables 82 and 83), the size sieving effects of UF should be explained. As earlier stated, UF is generally considered to separate solute and solvent via size-sieving. Thus a particle that has a size (generally expressed as a molecular weight) that will pass through the membrane's effective pore size will not be retained to any great detail in the concentrate. Particles larger than the nominal molecular weight cutoff will generally be retained in the concentrate. Thus the size of the particular parameter molecule will have a significant effect on the degree of membrane passage.

It has been reported that fulvic and humic acids are the main sources of color in natural waters have molecular weights of 100 to 1000 and 100,000 or greater, respectively. It has been also reported that humic and fulvic substances are common to raw water (Black and Christman, 1963) and present the predominant precursors of THM (Arguello et al., 1979). Given these sizes it is reasonable to expect that humic acids would be removed totally and a substantial proportion of fulvic acids would also be removed by the test membrane. Thus, from an UF size-sieving mode, it would be expected a majority of the color and THM precursors would be removed. In the case where diffusion is significant, the generally high molecular weight of the humic substances may significantly lower the solute diffusion constant and improve product quality.

Given the discussion above, the lack of an exact pattern in the color and THM values is more readily and plausibly explained. At low color and THM's product water values, an exact relationship is difficult to obtain because of the nature of the measurement of color and the analyses. However for the colors less than 1.5 cpu THM's does not exceed 28 μg/L in these studies. For these parameters it is reasonable to state that the main point is the very low color and THM's values obtained at all the tested combinations of feed pressure and product recovery rates.

Table A-16 and Figures 102, 103, and 104 exhibit the parameter removal percentages based on the raw water quality. The trends noted are consistent with the discussion above except that being removals, the trends are

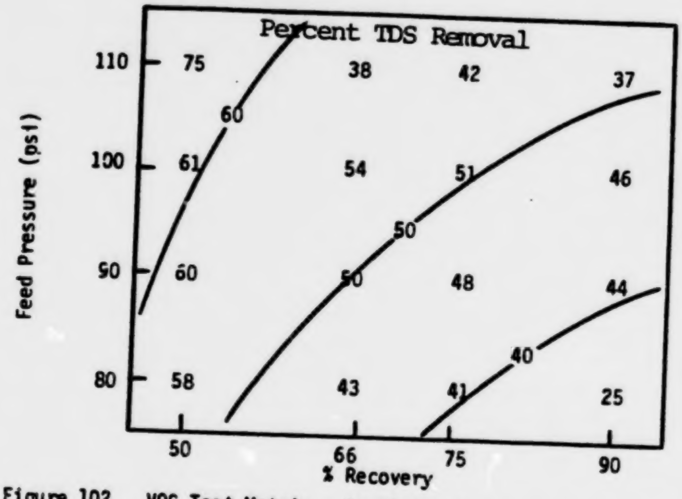


Figure 102 VOG Test Matrix - Isopleths for percent TDS removal at varying feed pressure and product recovery.

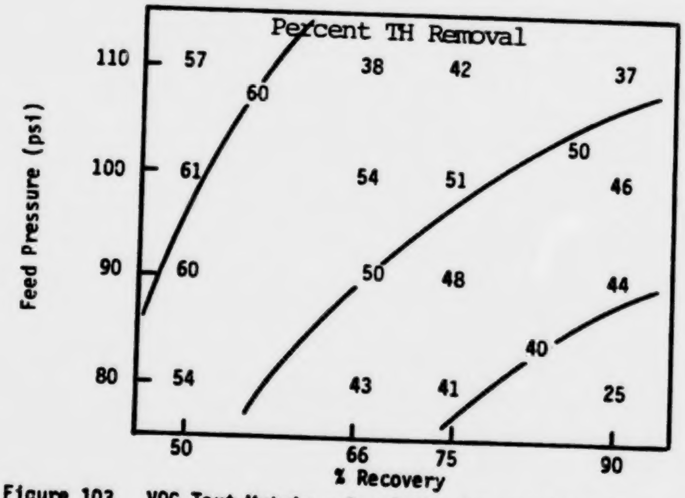


Figure 103 VOG Text Matrix - Isopleths for total hardness removal at varying feed pressure and product recovery.

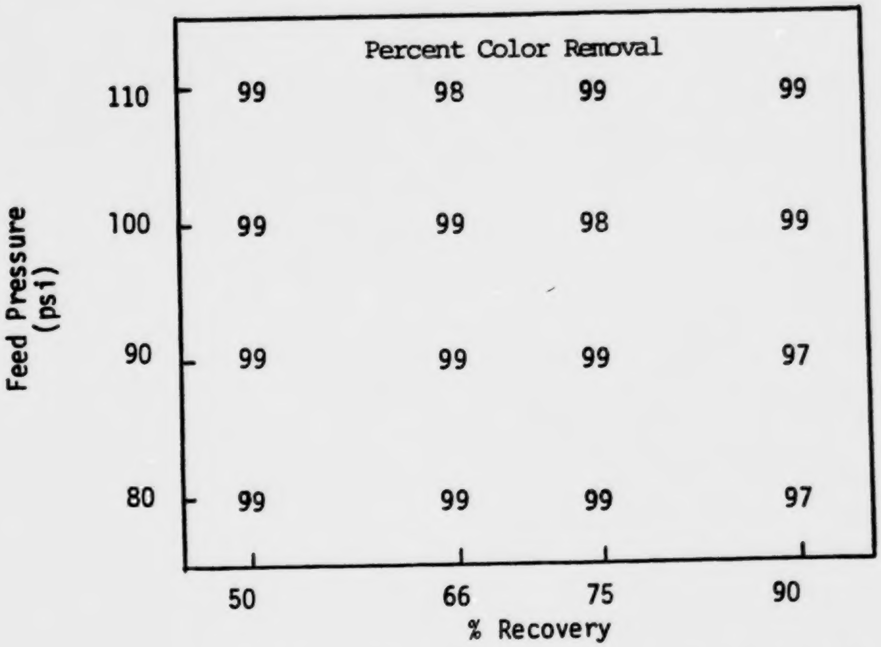


Figure 104 VOG Test Matrix Isopleth for percent color removal for varying feed pressure and product recovery

reversed. Thus as feed water solute concentration increase, the removals based on raw water will fall. It should be noted that color is removed almost completely.

At VOG prior to the full test matrix, the effect of varying feed pressure and varying product recovery rate was examined. During initial startup on December 28, 1984 the product recovery rate was altered from 50% to 90% at a constant feed pressure of 90 psi. The results may be seen in Table 86. Again the effect of increasing product recovery rate on product water solute concentration may be seen. It is also noteworthy in that DOC (not measured on 2-2-85) also displays an increase with increased product recovery rate. Also true color is once again reduced to a low and consistent value.

Table 87 gives the parameter removal percentage based on both raw and estimated feed water quality. While a specific feed sample was not taken, the use of raw and reject water quality and appropriate flow values allows the calculation of an value by difference. Examination of the raw-water based values again shows the expected drop in removal with increasing product recovery rate except for true color. Again the low levels provide a rather consistent removal (~95%) value. An examination of the feed-water based values show that DOC, TH and color all have relatively constant removal percentages. TDS has a more widely varied rate but except for the 90% product recovery rate value does not exhibit any substantial changes. It would appear that many forces afffecting product water quality are roughly balanced so as to preclude large deviations as the rate is varied. A comparison of Table 87 (at start-up) and Table 86 (approximately 200 hours run time) shows that with a few exceptions the removals based on raw water to be roughly comparible.

Further testing on January 18, 1985, also tends to confirm the previously discussed findings. Table 88 exhibits the product and feed water quality as product recovery rate was varied betwen 50% and 90% for both a pre- and post- membrane flush condition. As seen before the (product) inorganic parameters and DOC show increasing concentration with increased product recovery rate. Color THM's and TOXFP all exhibit low values with no consistent trends. Once again the THM's values are well below 0.1-mg/L. Inspection of Table 89 shows that the removal percentages correspond in trend to the previously discussed results. It should be noted that the membrane flushing did not significantly alter the removal percentages.

Also on the same date (1-18-85) the effect of increasing feed pressure at a constant product recovery rate of 75% (see Table 90) was tested. For this condition only very slight changes in water quality were noted for all parameters. This situation is reflected in the parameter removal percentages which generally vary only slightly. As stated before the effect of feed pressure is less definite and difficult to predict.

The long term testing at VOG was initiated on January 2, 1985, and terminated on March 3, 1985 after an elapsed operating time of 365 hours. Table A-17 and Figure 105 give a summary of the pilot plant operating data. From Figure 105 several trends in the operating parameters are apparent.

Table 86. VOG Initial Membrane Startup - Raw, Reject and Product Water Quality. (Testing on 12-28-84) Constant Feed Pressure = 90 psi

Sample	Product Recovery Rate (%)	Dissolved Organic Carbon (mg/L as C)	Total Dissolved Solids (mg/L)	Total Hardness (mg/L as $CaCO_3$)	True Color (cpu)
Raw	50	18.6	295	254	57
Reject	50	31.8	597	399	105
Product	50	0.8	73	109	3.0
Raw	66	15.9	268	256	57
Reject	66	47.1	718	563	165
Product	66	0.8	114	129	2.5
Raw	75	16.3	282	256	58
Reject	75	59.0	804	593	210
Product	75	0.9	116	154	3.0
Raw	80	18.1	283	256	58
Reject	80	62.3	800	601	230
Product	80	1.7	172	151	3.0
Raw	90	15.7	289	254	58
Reject	90	64.9	796	622	230
Product	90	1.9	229	149	3.0

Table 87. VOG Initial Membrane Startup - Parameter Removal Rates Based on Raw Water and Estimated Feed Water Quality

Basis for Comparison	Product Recovery Rate (%)	Dissolved Organic Carbon (%)	Total Dissolved Solids (%)	Total Hardness (%)	True Color (%)
Raw	50	96	75	57	95
Raw	66	95	57	50	96
Raw	75	94	59	40	95
Raw	80	91	39	41	95
Raw	90	88	21	41	95
Feed	50	96	75	57	95
Feed	66	96	82	64	96
Feed	75	89	79	64	98
Feed	80	96	71	67	98
Feed	90	95	58	66	98

Table 88. VOG Product and Feed Water Quality as a Function of Varying Product Recovery Rate (Feed Pressure = 90 psi) Testing performed on 1-18-85 both pre and post membrane flushing

Flush Condition Sample	Product Recovery Rate (%)	Dissolved Organic Carbon (mg/L as C)	True Color (cpu)	Total Dissolved Solids (mg/L)	Total Hardness (mg/L as $CaCO_3$)	Alkalinity (mg/L as $CaCO_3$)	Trihalomethane Formation Potential (µg/L)	Total Organic Halides Formation Potential (µg/L as Cl)
Pre-Feed	50	17.1	55	311	228	219	NA	NA
Pre-Product		1.3	1.5	181	94	98	31	29
Pre-Feed	66	25.8	80	413	281	269	NA	NA
Pre-Product		1.4	2.0	158	114	117	21	57
Pre-Feed	75	32.8	110	477	326	294	NA	NA
Pre-Product		1.6	2.5	194	128	134	20	44
Pre-Feed	90	36.0	120	528	342	319	NA	NA
Pre-Product		1.7	3.0	205	132	151	25	27
Post-Feed	50	17.8	55	343	226	218	NA	NA
Post-Product		1.2	2.0	145	94	92	20	26
Post-Feed	66	23.9	80	412	290	269	NA	NA
Post-Product		1.9	2.0	187	116	118	50	78
Post-Feed	75	31.4	105	441	324	298	NA	NA
Post-Product		1.4	2.5	203	125	130	28	24
Post-Feed	90	37.4	120	559	352	317	NA	NA
Post-Product		1.9	2.0	216	132	139	23	63

Table 89. VOG - Percent Parameter Removals as a Function of Varying Product Recovery Rate (Feed Pressure = 90 psi) Testing Performed on 1-18-85 Both Pre- and Post-Membrane Flushing

Flush Condition	Basis of Calculation	Product Recovery Rate (%)	Dissolved Organic Carbon (%)	True Color (%)	Total Dissolved Solids (%)	Total Hardness (%)	Alkalinity (%)
Pre	Raw	50	92	97	42	59	55
Pre	Raw	66	92	96	49	50	47
Pre	Raw	75	91	95	38	44	39
Pre	Raw	90	90	95	34	42	31
Post	Raw	50	93	96	58	58	58
Post	Raw	66	89	96	45	49	46
Post	Raw	75	92	95	41	45	40
Post	Raw	90	89	96	37	42	36
Pre	Feed	50	92	97	42	59	55
Pre	Feed	66	95	98	62	59	57
Pre	Feed	75	95	98	59	61	54
Pre	Feed	90	95	98	61	61	53
Post	Feed	50	93	96	58	58	58
Post	Feed	66	92	98	55	60	56
Post	Feed	75	96	98	54	61	56
Post	Feed	90	95	98	61	63	56

Table 90. VOG - Product, Raw, and Feed Water Quality and Percent Parameter Removals as a Function of Varying Feed Pressure (Product Recovery Rate = 75%) Testing Performed on 1-18-85 on a Post-Membrane Flushing Condition.

Sample	Feed Pressure (psi)	Dissolved Organic Carbon (mg/L as C)	True Color (cpu)	Total Dissolved Solids (mg/L)	Total Hardness (mg/L as $CaCO_3$)	Alkalinity (mg/L as $CaCO_3$)	Trihalomethane Formation Potential (µg/L)	Total Organic Halides Formation Potential (µg/L as Cl)
Feed	80	32.8	100	382	317	288	NA	NA
Product	80	1.9	2.0	198	125	132	17	24
% Removal Raw		89	96	42	45	39	NA	NA
% Removal Feed		94	98	48	61	54	NA	NA
Feed	100	34.1	105	515	326	300	NA	NA
Product	100	1.3	2.0	209	125	130	17	33
%Removal Raw		93	96	39	45	40	NA	NA
%Removal Feed		96	98	59	62	57	NA	NA
Feed	110	37.5	110	513	333	308	NA	NA
Product	110	NA	2.0	210	123	128	16	21
% Removal Raw		NA	96	39	46	41	NA	NA
% Removal Feed		NA	98	59	63	58	NA	NA
Raw		17.8	55	343	226	218	NA	NA

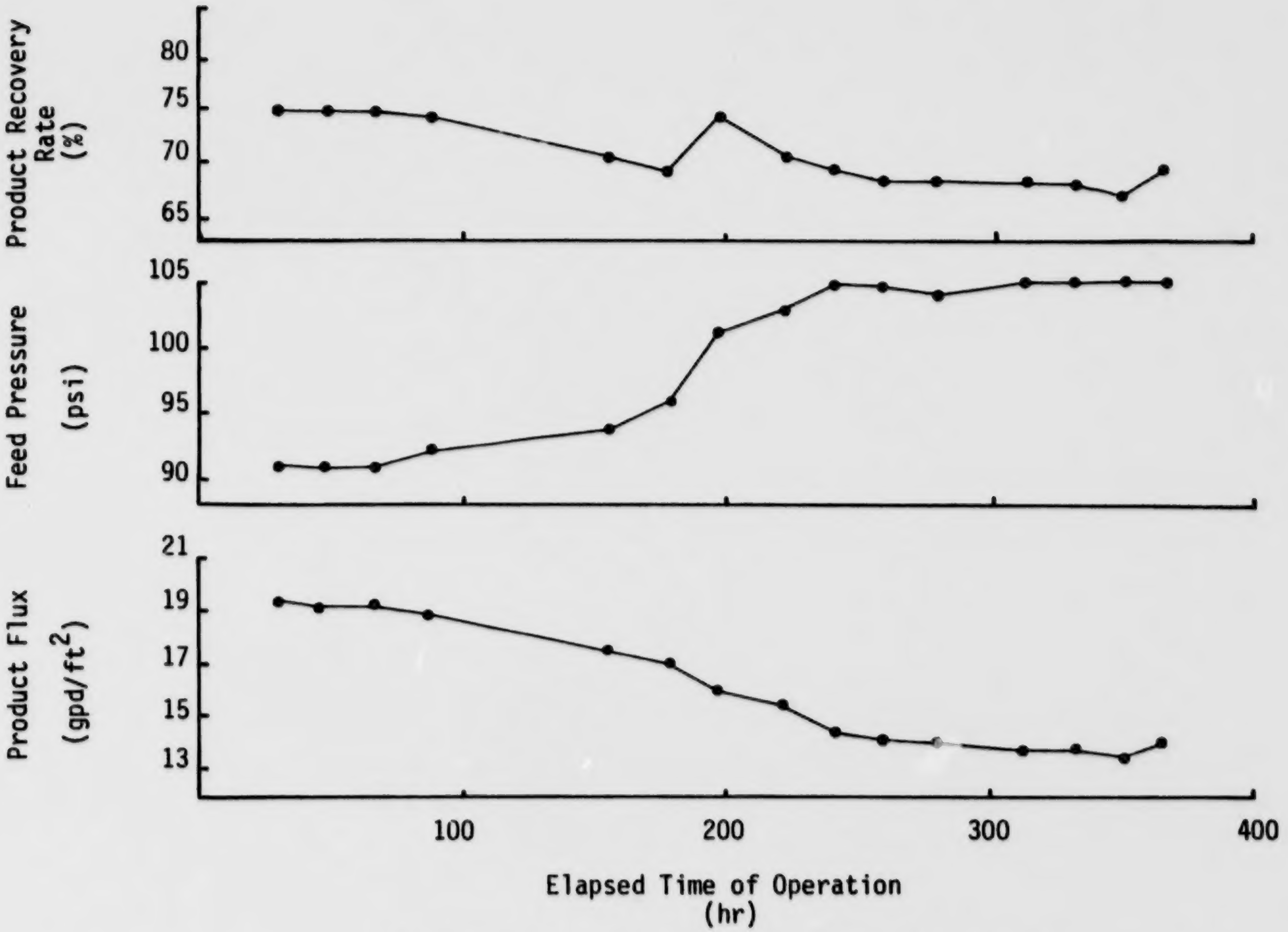


Figure 105 VOG Pilot Plant Operation-Product Recovery Rate, Feed Pressure and Product Flux versus Time of Operation.

Throughout the duration of the test, a steady decline in product flux was noted. The lowest observed flux value of 13.4 gpd/ft^2 represented a 33% decline from the maximum value of 20.0 gpd/ft^2 which is the Filmtec design flux. This decline is further accentuated by the fact the operating feed pressure was approximately 15 psi higher during the low flux value as compared to the earlier high value. Such a flux loss was attributed to a clogging or "blinding" of the membrane. The flux returned to 20 gpd/ft^2 following chemical cleaning at AID. This flux decline was probably due to biological growth promoted by an 8 hr/day operation and/or the organics in the raw water.

Concurrent with the noted flux decline was a steady increase in feed pressure and a decline in percent product recovery. Such phenomenon are attributable to membrane blinding. As the membrane blinds, an increased feed pressure is necessary for solvent transport. In conjunction the percent of product recovery declines since the product flow is decreasing while reject water flows remain constant. It should be noted that the jump in feed pressure immediately after the 180 hour mark was a deliberate operational change.

To aid the discussion of the various water quality parameters, several common tables have been provided. Tables A-18, A-19, A-20, and A-21 provide a data summary of the raw, feed, reject and product water streams respectively. Table A-22 gives the TOXFP and THM's data. Tables A-23 and A-24 provide the percent parameter removals in the product water based on feed and raw water values respectively. Table 91 gives the correlation coefficients between the product water parameters and the feed and raw water parameters while Table 92 gives the correlation coefficients between the product water parameters. It is generally recognized that DOC and color can give a relative indication of expected THM's and TOXFP values values. To aid the discussion of the THM's and TOXFP results, the DOC and color parameters shall be discussed first.

Figure 106 gives a graphical representation of the raw, feed and product water concentrations on a time of operation basis. As can be seen the product water DOC has a fairly constant value in the 1.5-mg/L to 2.5-mg/L range with an average value of 1.9-mg/L. Feed water values during the testing period ranged from 29.8-mg/L to 33.2-mg/L with an average value of 31.7-mg/L. Similar numerical ranges were found for the raw water values (13.8-mg/L to 17.4-mg/L) with an average value of 15.9-mg/L. The effectiveness of the test membrane in removing DOC can be seen in the percent removal values (see Figure 107). The observed product water values represented a removal range of 92% to 95% based on feed water and 83% to 91% based on raw water. The average removal was 94% and 88% based on feed and raw water respectively.

Examination of Table 91 shows that the product water DOC value is relatively weakly correlated with both the feed and raw water DOC value. This fact would tend to support molecular size seiving as the principle mechanism affecting organic separation. If the passage of DOC was concentration-dependent, then a relatively good correlation between product and feed values would be expected. However it should be further recognized

TABLE 91. VOG Linear Correlation Coefficients Product Water Parameters *vs* Feed and Raw Water Parameters

Product Water	Feed Water					Raw Water				
	DOC	Color	Alakalinity	T. Hardness	TDS	DOC	Color	Alakalinity	T. Hardness	TDS
DOC	0.35	-0.22	0.00	-.13	-0.22	0.24	-0.12	0.14	0.28	-0.10
Color	0.12	0.06	-0.06	0.48	0.11	0.01	-0.44	0.03	0.62	0.24
Alkalinity	0.11	0.45	0.70	0.73	0.62	-0.40	-0.38	0.81	0.54	0.15
T. Hardness	0.11	-0.03	-0.06	0.47	-0.08	0.15	-0.05	-0.06	00.76	0.12
TDS	-0.26	0.78	0.82	0.64	0.81	-0.66	-0.46	0.85	0.24	0.48
THMFP	0.03	0.23	-0.04	0.34	0.22	-0.29	0.30	-0.07	0.45	-0.37
TOXFP	-0.04	0.55	0.48	0.70	0.52	-0.47	-0.22	0.52	0.57	0.15

TABLE 92. VOG. Linear Correlation Coefficients - Product Water Parameters

Parameter	DOC	Color	Alkalinity	T. Hardness	TDS	THMFP	TOXFP
DOC	1.00	0.56	0.45	0.53	-0.19	0.10	0.21
Color	0.56	1.00	0.37	0.74	0.00	0.22	0.43
Alkalinity	0.45	0.36	1.00	0.30	0.62	0.07	0.60
T. Hardness	0.53	0.74	0.30	1.00	-0.19	0.40	0.43
TDS	-0.19	0.00	0.62	-0.19	1.00	-0.13	0.57
THMFP	0.10	0.22	0.07	0.40	-0.13	1.00	0.49
TOXFP	0.21	0.43	0.60	0.43	0.57	0.49	1.00

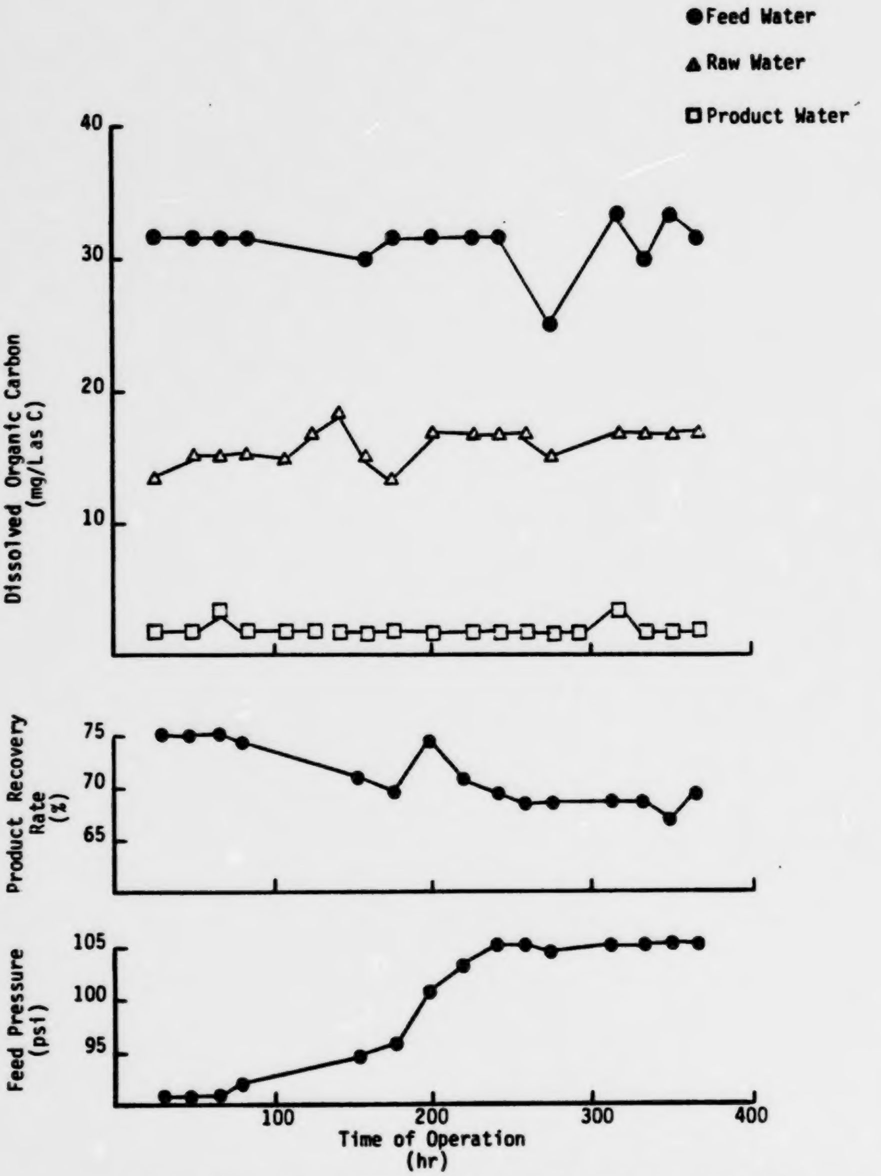


Figure 106 VOG Membrane Pilot Plant Operation - Dissolved Organic carbon versus time of operation, product recovery rate and feed pressure.

that the relative size fractionation in the feed water and the effects of a concentration gradient on the solvent (water) flux will further complicate the system. Thus, while the product water DOC value is not easily correlated to the feed water value, it can be stated that given the present set of conditions, a very significant reduction in DOC occurred through the use of the test membrane. This data is more descriptive of a UF process than a RO process.

Figure 108 gives the plots of the raw, feed, and product water true color values versus the time of operation. As can be seen the product water values are quite low (1.0 cpu to 4.0 cpu) with an average of 2.0 cpu. It should be noted that the 4.0 cpu value was taken at the clear well and could be an artifact of algae growth in the aerator. Subsequent to this sampling date, the product water sample was taken from the product water collection manifold prior to the aerator. Without this value the maximum observed value was 3.0 cpu. The feed water values ranged from 90 cpu to 120 cpu with an average of 105 cpu. The corresponding raw water value range was 55 cpu to 61 cpu with an average of 58 cpu.

Tables A-23 and A-24 and Figures 107 and 108 confirm the expected excellent removals of color. Removal is 97% or greater based on feed water values and 93% or better based on raw water values. Given the nominal molecular size cut-off of 400 MW for the test membrane and the fact that the majority of fulvic and humic acids are generally present in molecular weights greater than 500, the excellent removals are expected. Such a low color value eliminates the need for chlorine as a bleaching agent.

Inspection of Table 91 shows that product water color has essentially no correlation with feed color and only weak correlation with raw color. Again it would appear that color is removed to a relatively uniform value due to size-seiving as would be expected in UF. It should be noted that this situation is similar to the situation discussed for DOC.

The data for THM's are presented in Table A-22 and Figure 109 and represent product water values chlorinated under laboratory conditions. In all cases the 7-mg/L Cl_2 initial dose was sufficient to maintain a residual greater than 0.2-mg/L for the 96 hour duration. In several instances the residual was just at the limit. As can be seen, the THM's are comfortably below the 100 µg/L MCL in all cases. In several cases the values are quite low (below 20 µg/L). While care was taken in all phases of the chlorination, pH adjustment, incubation, quenching and eventual analysis, the complex nature of the process will, in fact, induce some variability. However, such a process could also lead to over-chlorination as well. Thus it is important to note that an overall average of 27 µg/L was obtained during the duration of the test. Such values would be expected from the low DOC and color values obtained in the product water.

Inspection of Table 92 shows that the product THMFP value is only weakly linearly correlated with either feed DOC or color or raw DOC or color. Further inspection of Table 92, which gives correlations between the product water parameters, shows that a linear correlation between THMFP and either product water DOC or color to be almost non-existent. While

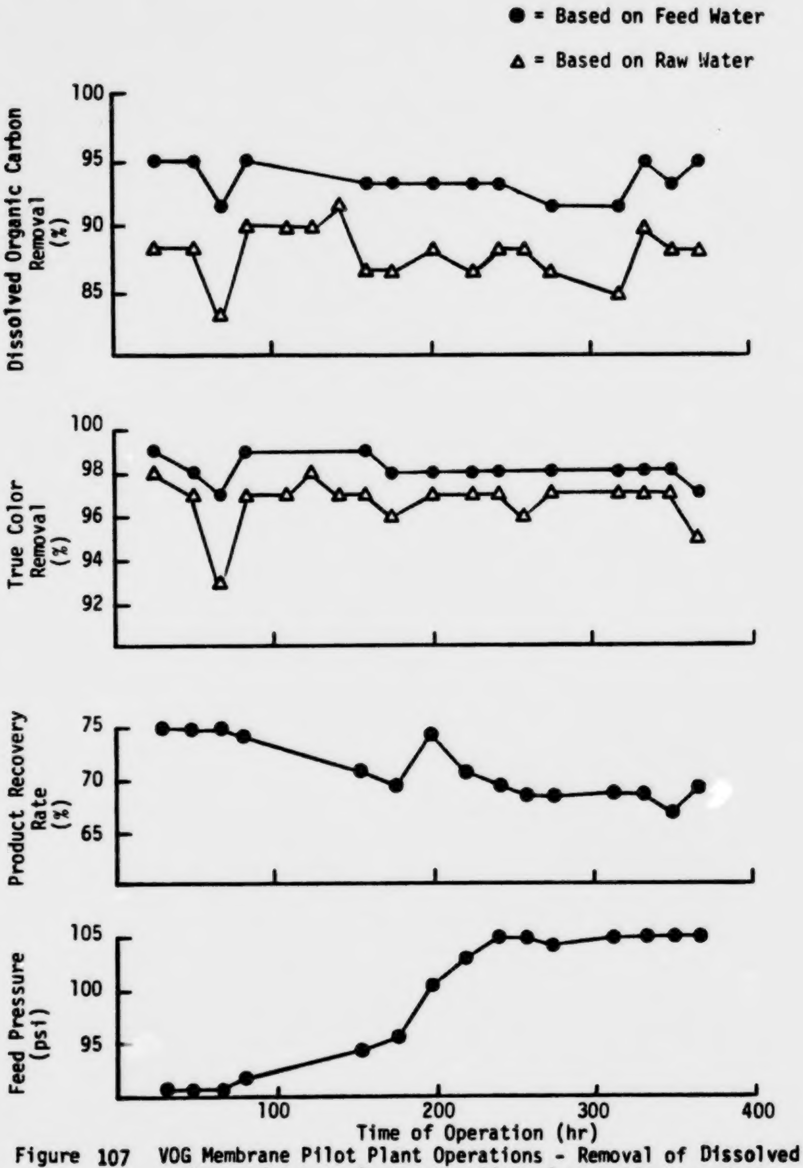


Figure 107 VOG Membrane Pilot Plant Operations - Removal of Dissolved Organic carbon and true color versus time of operation, product recovery rate and feed pressure.

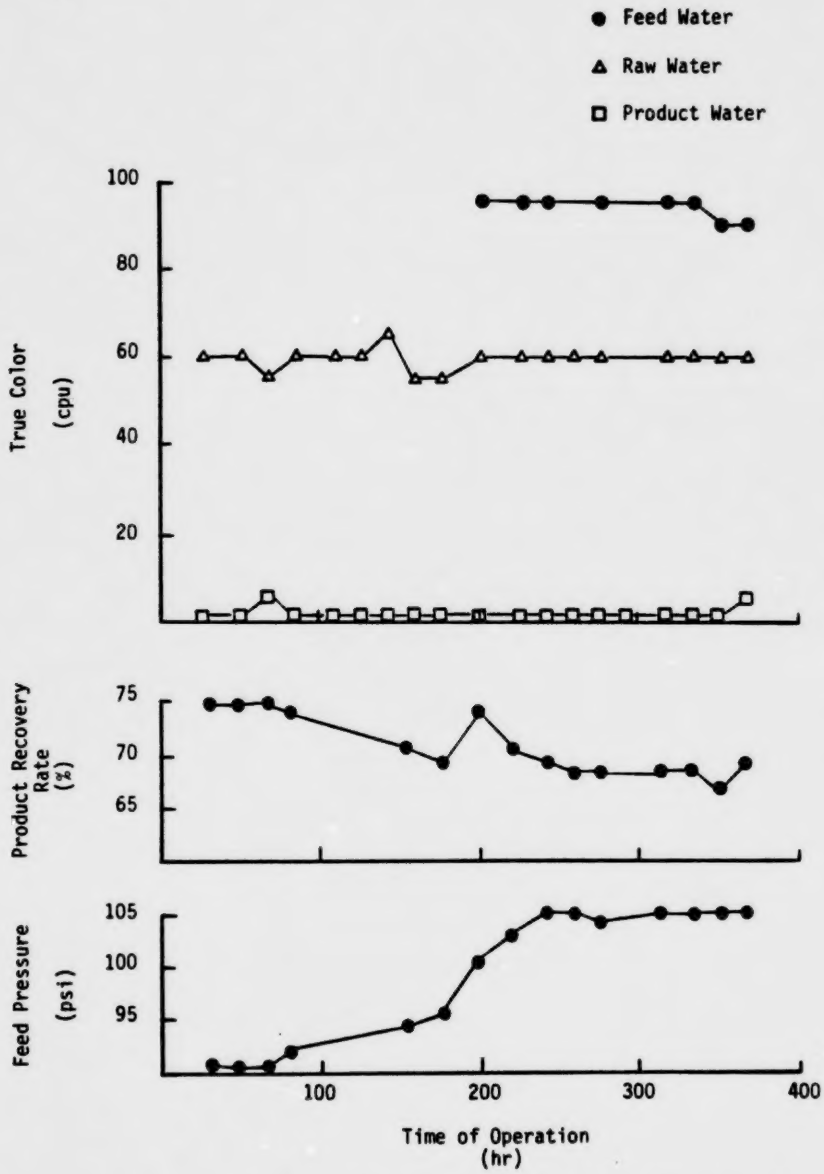


Figure 108 VOG Membrane Pilot Plant Operation. True color versus time of operation, product recovery rate and feed pressure.

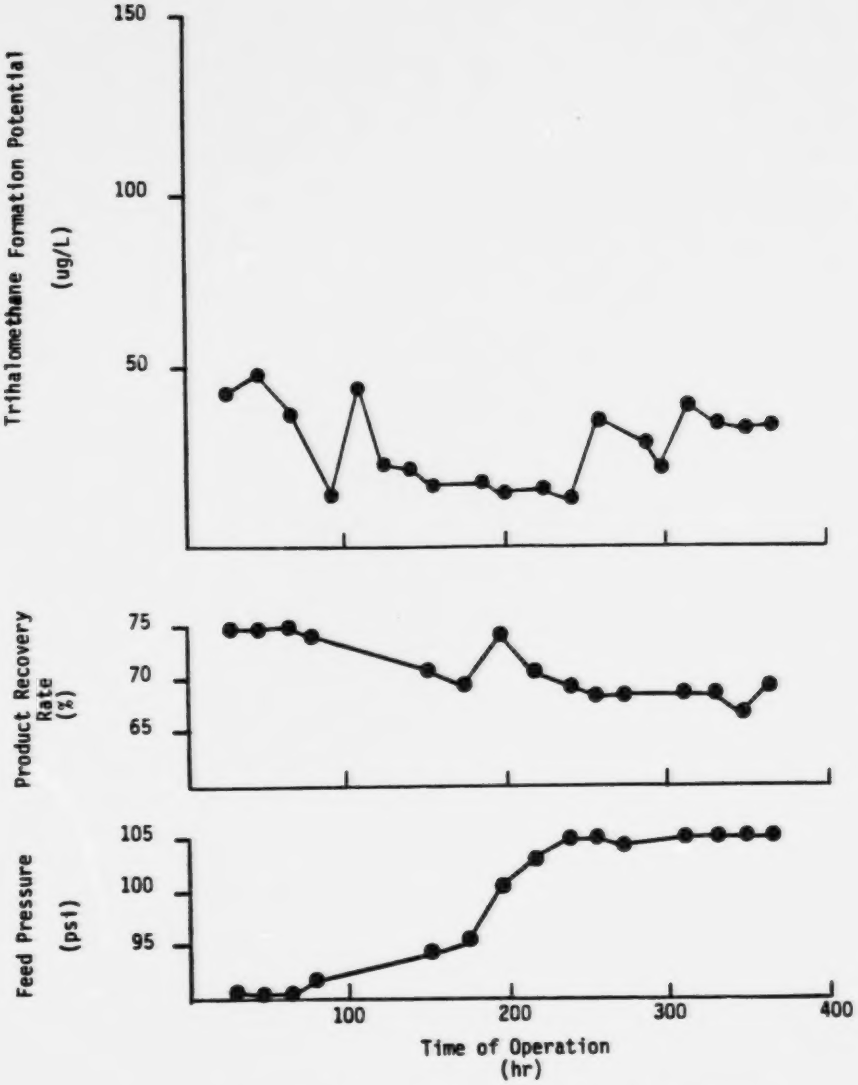


Figure 109 VOG Membrane Pilot Plant Operation. Trihalomethane formation potential versus time of operation, product recovery rate and feed pressure.

intuitively it would be expected that DOC and color would be related to the THM's value, at the low values observed random variations from the chlorination and analysis procedure could obscure the trends.

The data for TOXFP are contained in Table A-22 and Figure 110. As can be seen the data show a wider range of values (15 μg/L to 120 μg/L) for TOXFP than for THM's (13 μg/L to 47 μg/L). Much of the discussion involving the THM's values would be applicable to the TOXFP value. At the current time a MCL does not exist for TOXFP thus a discussion of meeting a MCL is not appropriate. Inspection of Tables 91 and 92 shows that TOXFP is not substantially linearly correlated with either feed or raw water DOC and color. Also TOXFP is not substantially linearly correlated with THM's in the product water. This may be confirmed by the data contained in Table A-22 that shows the ratio between TOXFP and THM's. As is noted the ratio varies between 0.52 and 5.07, a nearly ten-fold difference. The average ratio is 1.93 using the units normally reported. Again the wide range must be partially due to analytical procedures especially given the relatively low values.

The plots of raw, feed and product water values for TDS versus time of operation may be seen in Figure 111. The range found in the product water is 172-mg/L to 215-mg/L with an average value of 195-mg/L. The feed water ranged from 44-mg/L to 659-mg/L with an average of 510-mg/L. The comparable raw water range was 321-mg/L to 416-mg/L with an average of 367-mg/L. The product water values are well-below the recommended limit of 500-mg/L. From Table A-23 and A-24 and Figure 112 it can be seen that the percent removal ranges from 57% to 67% and 39% to 53 % based on feed and raw waters respectively. While this removal figure is relatively lower than values usually experienced for a RO operation, it is significantly higher than the removal values typically found for UF operations.

Examination of Table 91 shows that the product water TDS concentration is relatively well-correlated with the feed water value (+0.82). The species that constitute the majority of the TDS value are generally lower molecular weight inorganic compounds. Such lower molecular weight substances could pass the membrane either through pore passage or diffusion through the membrane. In either case, a higher concentration would either bring a greater number of particles to the pore openings or increase the concentration gradient for diffusion. The situation is complicated and removed from the ideal in that, due to the organic layer, metal ions can be either complexed and/or chelated. Such chemical phenomenon would hinder the passage of these metal species. Also in the concentrated layers in the membrane surface area, the possibility of precipitation or polymerization would increase, again slowing the transport of the metal species.

The raw, feed and product water total hardness values are plotted against time of operation in Figure 113. The product water average was 142-mg/l as $CaCO_3$ with a relatively wide range of 120-mg/L as $CaCO_3$ to 179-mg/L as $CaCO_3$. The feed water average was 342-mg/L as $CaCO_3$ with a range from 323 to 375-mg/L as $CaCO_3$. The raw water values ranged from 228-mg/L as $CaCO_3$ to 284-mg/L as $CaCO_3$. The product water value is higher than the current plant average of approximately 80-mg/L as $CaCO_3$. However

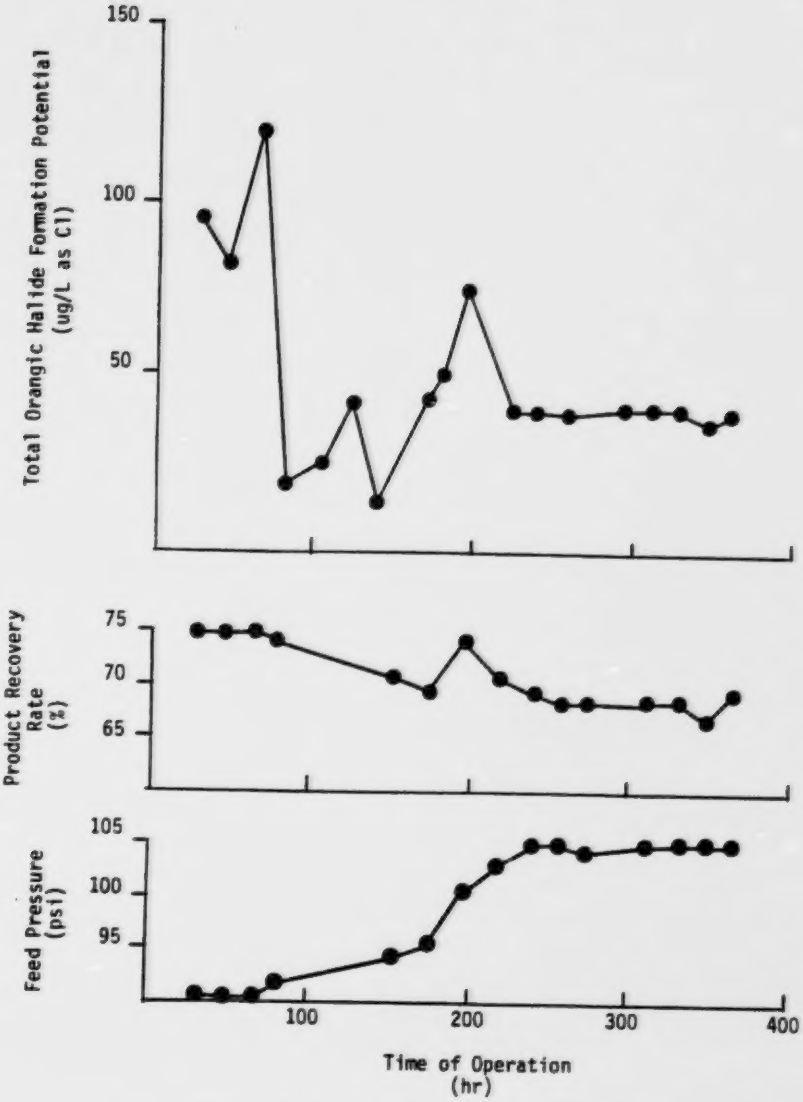


Figure 110 VOG Membrane Pilot Plant Operation. Total organic halide versus time of operation, product recovery rate and feed pressure.

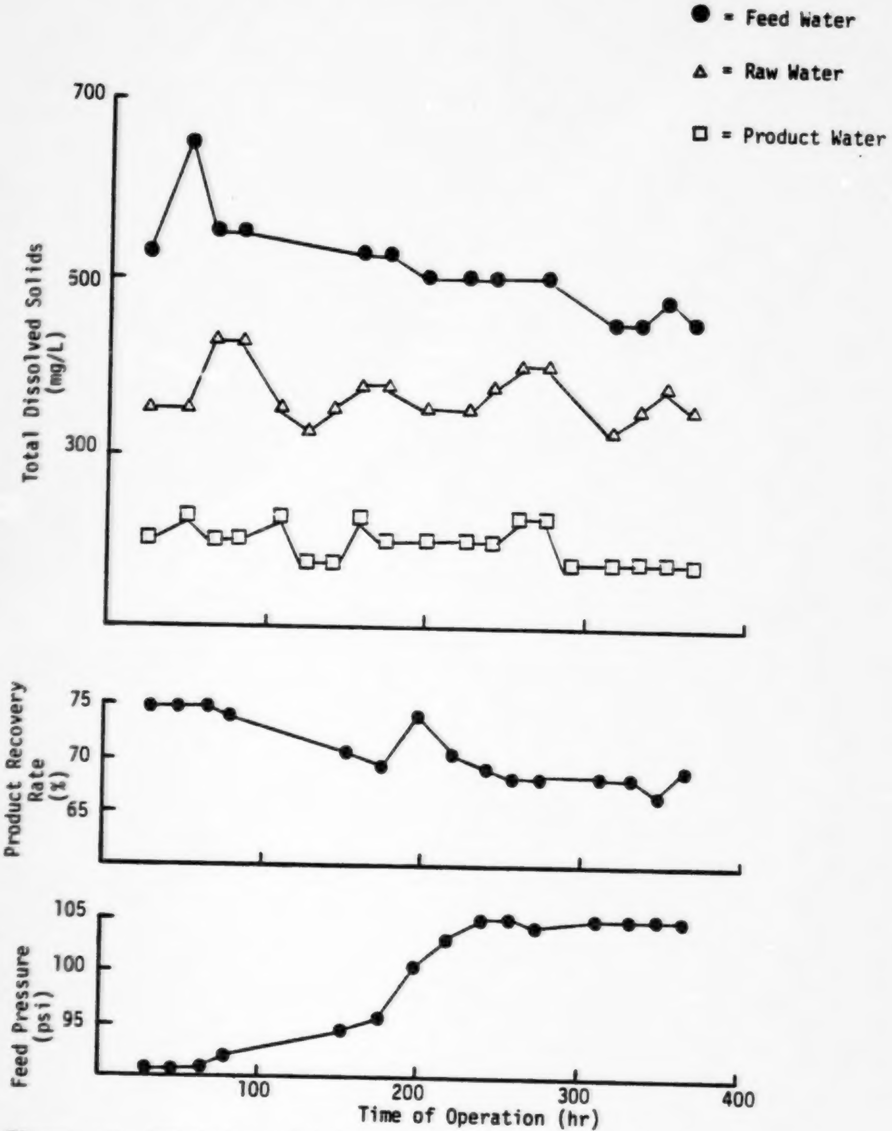


Figure 111 VOG Membrane Pilot Plant Operation - Total dissolved solids versus time of operation, product recovery rate and feed pressure.

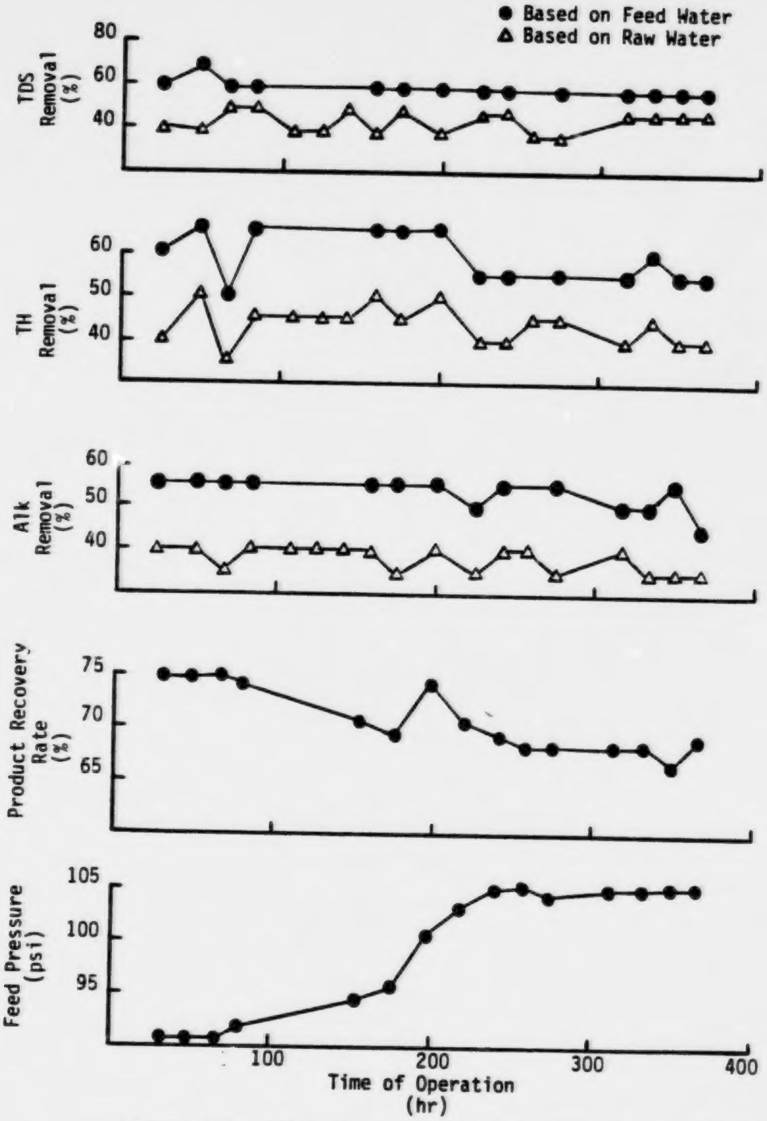


Figure 112 VOG Membrane Pilot Plant Operation - Removal of total dissolved solids, total hardness and alkalinity versus time of operation, product recovery rate and feed pressure.

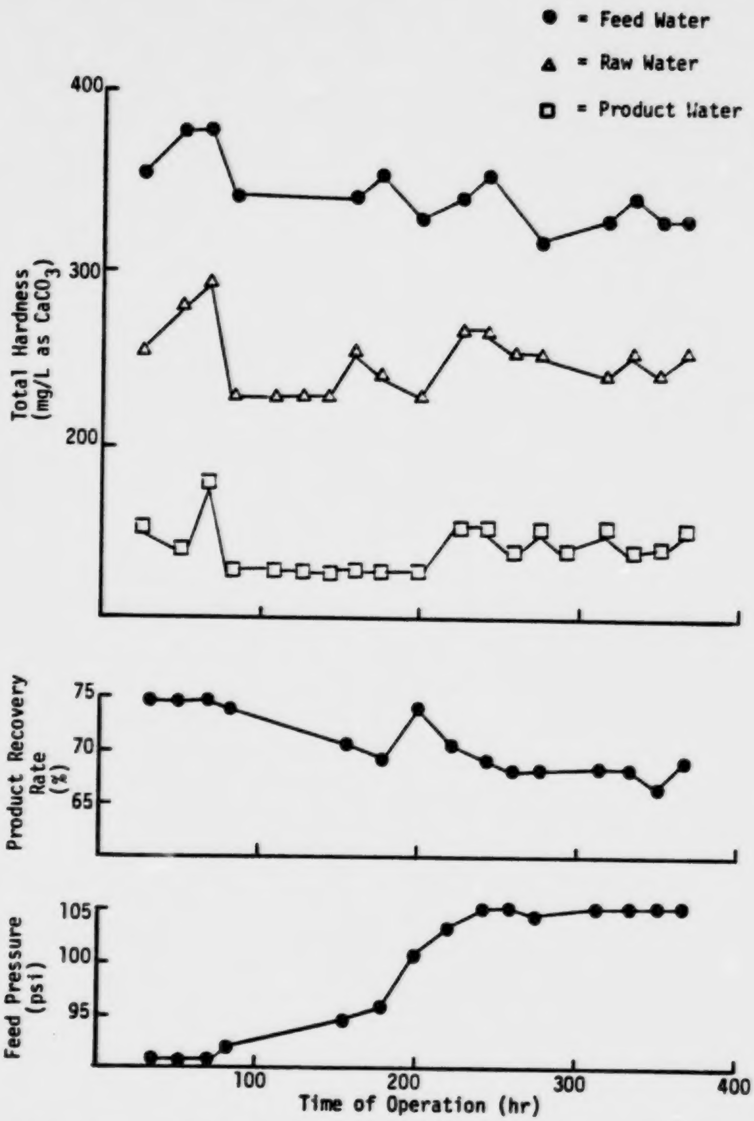


Figure 113 VOG Membrane Pilot Plant Operation - Total hardness versus time of operation, product recovery rate and feed pressure.

the hardness is rather typical of values found for many communities that draw their supply from Florida groundwaters.

The removals for TH in the product water (see Figure 112) are comparable, albeit slightly lower, to the values found for TDS (52% to 63% and 37% to 49%, compared to feed and raw water respectively). However the correlations between the product and feed water are substantially less than TDS. One factor could be the complexation and chelation reactions that can occur to a greater degree for polyvalent cations than for monovalent cations. However the relative effects of chelation on solute transport is complex and difficult to define. Also the use of an anti-scalent chemical would tend to complicate the behavior of the multi-charged ions.

The data for the various alkalinity values are plotted in Figure 114. The range of product water values is rather narrow (127-mg/L as $CaCO_3$ to 143-mg/L as $CaCO_3$) with an average of 135-mg/L as $CaCO_3$. The raw and feed ranges were 202-mg/L as $CaCO_3$ to 234-mg/L $CaCO_3$ and 241-mg/L as $CaCO_3$ to 322-mg/L as $CaCO_3$ respectively. The percent removals in the product waters (see Figure 112) were comparable to the values seen for TH (47% to 57% and 36% to 41% based on feed and raw values, respectively).

An examination of Table 91 shows that the product water alkalinity is moderately correlated to the feed water value (+0.70). However as in the case of TH and TDS the effect of chemical reactions at the membrane surface and the effect of the concentration gradient are unevaluated factors. The prediction of alkalinity values are not exact but do reflect a general 38% reduction based on feed.

6.4.7.3 Summary

To summarize the pilot plant operation at VOG, several salient points can be made. The main object of the testing was to establish the capability of the system to produce a water that could meet the THM standard. In this respect the test was quite successful with all THM's values well below the 0.10-mg/L standard. The product water also exhibited low DOC and low color values. The TDS values were well below the recommend level of 500-mg/L and the TH was in the acceptable range. Subsequent to chlorination (or other alternate disinfectants) the water would be suitable for potable purposes.

From the flux measurements it would appear that cleaning would be needed periodically to maintain flux at appropriate levels. The intermittent daily operation at VOG could promote bacterial growth which could decrease flux. As will be seen in the AID discussion, cleaning restored the flux to the original value. It should also be pointed out that the exact prediction of product water quality at any certain time would be difficult. However it is possible to state ranges of values (some of which are relatively narrow) from certain feed water qualities and operational conditions.

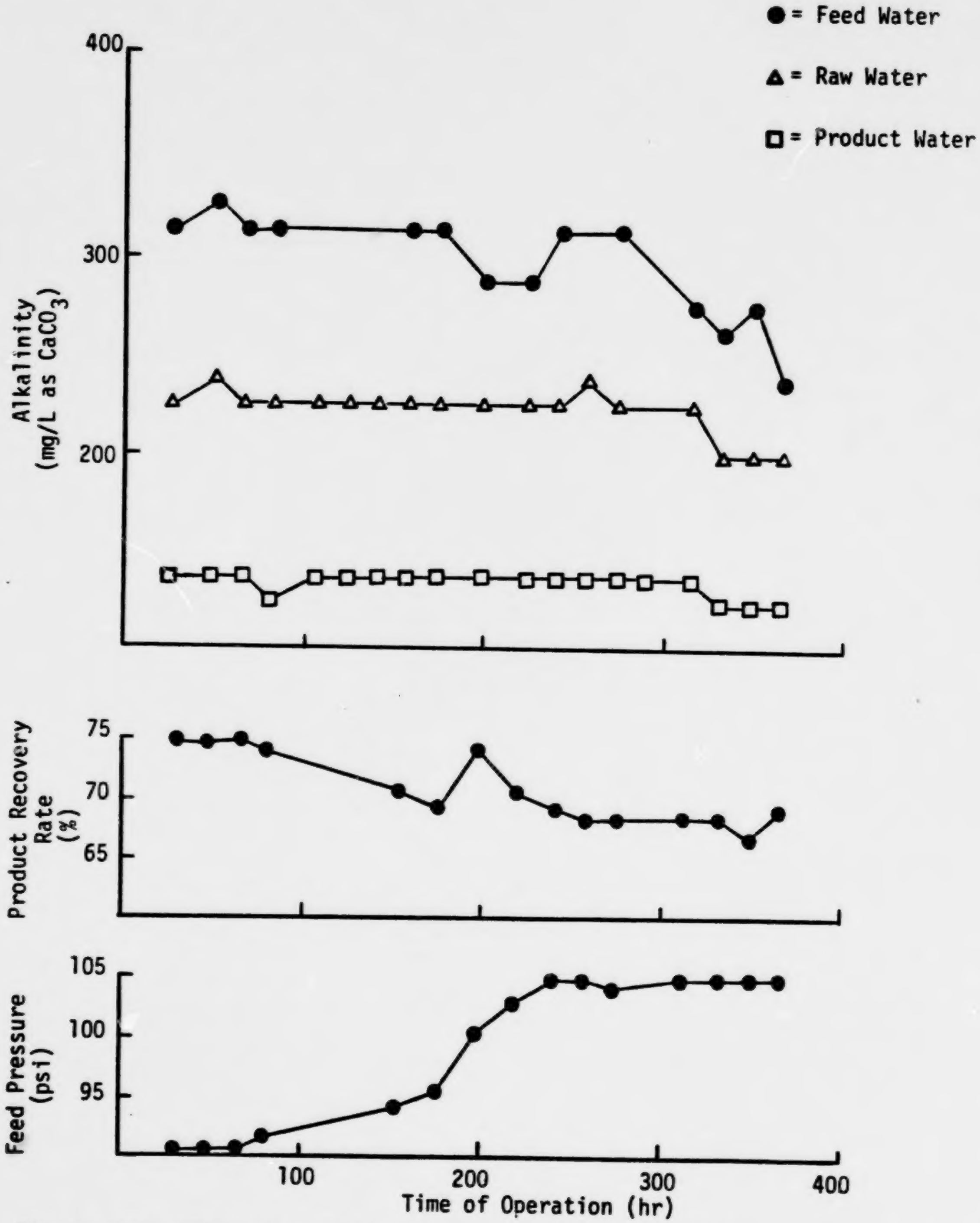


Figure 114 VOG Membrane Pilot Plant Operation - Alkalinity versus time of operation, product recovery rate and feed pressure.

6.4.8 AID Membrane Tests

6.4.8.1 Optimum Membrane Testing

The data for the BW30, N-50, and U90-G10 membrane bench-scale testing at AID may be found in Tables 93, 94, and 95. Table 93 provides the basic water quality data including percent removals on a raw water basis. Tables 94 and 95 provide the THM formation and chlorine residual data respectively for incubation periods up to 192 hours.

The relative performance of the three membranes was similar to that found at VOG. At AID the BW30 membrane produced a water with the characteristics of an RO product water. The DOC and color values of 0.62-mg/L and 1 cpu respectively represented a reduction of 96% and 97% from the raw water values respectively. The excellent organic removals are further illustrated by the low THM's values (23 μg/L and 27 μg/L) found for both a 5 mg/L and 10 mg/L initial chlorine dose. The alkalinity and total hardness were each reduced by 96% from the raw water. The product water alkalinity and total hardness values of 14 mg/L as $CaCO_3$ would indicate a very soft water with virtually no buffering capacity. It is interesting to note that the organic removals are less at the AID site as compared to the VOG site while the inorganic removals are higher; however, the differences are slight.

Further inspection of Tables 93, 94, and 95 shows that the N50 membrane also produces a suitable water for meeting the THM MCL. The product water DOC value of 1.41 mg/L represents a 90% removal based on the raw water while the 1 cpu color value represents a 97% reduction. Chlorination of the product water at the initial doses of 5-mg/L and 10-mg/L produced THM's values of 44 μg/L and 64 μg/L respectively after an incubation period of 192 hours. These THM values are comfortably under the current MCL. The product water alkalinity and total hardness values of 126-mg/L as $CaCO_3$ and 97 mg/L as $CaCO_3$ represent reductions of 61% and 70% based on raw water values respectively. The product water would be classified as a moderately hard water with good buffering capacity. Once again the reductions in organic constituents is slightly less than at VOG while the inorganic parameter removal is slightly greater.

Inspection of Tables 94 and 95 shows that the product water from the U90-G10 membrane would be unsuitable for meeting the THM MCL under the present conditions. Also the total hardness present in the product water would make the water very "hard" and not desirable as a potable water. This membrane was then eliminated as a possible choice for the pilot plant operation.

As in the case of VOG, two membranes are suitable for meeting the THM standard. Thus the decision on a membrane to be used for pilot plant studies was made on other considerations. Again, the N-50 membrane was chosen as the preferred membrane because of cost and quality. The N-50 has an operating pressure that is at a minimum approximately 150 psi less than the BW30. For a 1.5 MGD installation at AID, the yearly power savings would be approximately 79,500 kw-hr. The N-50 product water is more

Table 93. AID Optimum Membrane Testing Data and Product Water Parameter Removals

Sample and Membrane	pH	DOC (mg/L)	Color (cpu)	Alkalinity (mg/L $CaCO_3$)	CaH (mg/L $CaCO_3$)	TH (mg/L $CaCO_3$)	Turbidity (NTU)
Acme U90-G10	8.1	7.42	14	300	288	302	0.13
Raw	7.7	12.46	35	342	328	348	0.26
% Removal	--	40	60	12	12	13	50
Acme BW30	7.9	0.62	1	14	10	14	0.13
Raw	8.2	13.93	32	311	309	322	0.25
% Removal	--	96	97	96	97	96	48
Acme FT-50	7.3	1.41	1	126	93	97	0.09
Raw	7.3	14.64	35	325	304	319	0.25
% Removal	--	90	97	61	69	70	64

Table 94. AID Optimum Membrane Testing - THMFP Formation Curve (all values in µg/L).

Site	Membrane	Initial Cl_2 Dose (mg/L Cl_2)	Time (hrs.) 1	3	24	48	72	96	192
AID	BW30	5	11	11	14	12	15	34	23
AID	BW30	10	11	11	12	12	16	59	27
AID	FT-50	5	11	16	14	15	48	48	64
AID	FT-50	10	11	17	19	22	46	65	44
AID	U90-G10	5	45	75	62	48	70	97	60
AID	U90-G10	10	83	108	160	302	86	375	371

Table 95. AID Optimum Membrane Testing - Chlorine Residual Curve Data (all values in mg/L-Cl_2).

Site	Membrane	Initial Cl_2 Dose (mg/L Cl_2)	Time (hrs.) 1	3	24	48	72	96	192
AID	BW30	5	NA	3.8	3.4	3.3	3.1	3.5	2.7
AID	BW30	10	8.9	8.2	8.1	8.3	5.6	4.5	1.9
AID	FT-50	5	1.8	1.8	1.5	1.4	1.4	1.3	0.3
AID	FT-50	10	9.3	9.4	7.4	6.9	6.9	5.8	3.8
AID	U90-G10	5	0.72	0	--	--	--	--	--
AID	U90-G10	10	6.9	5.6	4.2	3.4	2.7	1.4	1.0

suitable as a potable water due to its buffering capacity and moderate hardness. The BW30 product water would have virtually no buffering capacity and would be corrosive upon chlorination.

6.4.8.2 Secondary Membrane Testing

As earlier discussed, four additional membranes were tested at AID with the product water tested for DOC, color, and TH and then chlorinated for THM's and chlorine residual measurements. From the stated molecular cutoff weights, it was not expected that these membranes would produce a suitable product water. This assumption is confirmed by the data given in Tables 96, 97, and 98. All the membranes exhibit only small reductions in parameter levels from the raw water. Also, generally large (20 mg/L) chlorine doses are required for the 96-hour residuals and the resultant THM's values greatly exceed the MCL.

The secondary membrane testing was, however, useful in determining the relative distribution of the tested parameters on a molecular size basis. Table 99 gives both the percentage of parameter removed from the raw and percent of parameter still present in the product water for DOC, color, and TH. For THM's, the values are given and the percent removal compared to the highest observed values. The secondary membrane values were derived from the 50 psi feed pressure values while the other membranes' values were from the previously discussed samples. As can be seen from the Table 99, the percent remaining in the product generally increases as the molecular weight cutoff increases. In the case of THM's, the percent removal from the highest value increases with decreasing molecular weight cutoff values.

The U90-GSO membrane which has a stated molecular weight cutoff value of 20,000 (same as the Osmo PT-2), regularly exhibits a product water quality that is inconsistent with the stated cutoff value. In particular the percent removals based on raw water are much higher than the PT-2 membrane and also slightly higher than the UFP-4040 (10,000 MW). At this time, no readily apparent explanation can be ascribed to this phenomenon. One possibility could be that the membrane through manufacturing variableness, has a "tighter" than expected pore size. Another possibility is that the particular chemical makeup of the membrane is allowing some unknown chemical reactions that aid in retention.

6.4.8.3 Pilot Plant Studies

At AID a feed pressure product recovery rate test matrix was performed on March 13 and 14, 1985 during initial operation. The feed pressure and product recovery rate was altered in a manner similar to that at VOG. The results may be seen in Tables A-25, 100, 101, 102, 103, and 104. For the inorganic parameters TDS and TH and the organic parameter DOC, an increase is seen with increasing product recovery rate at a constant feed pressure. Also for these parameters, relative slight and inconsistent changes in product water quality are observed with increasing feed pressure. Color was once again present at a low value and displayed no consistent trends. Examination of the THM's data (Table 101) shows that, at the three highest feed pressures (90 psi, 100 psi and 110 psi), an increase in THM's is seen

Table 96. Secondary Membrane Testing Water Quality Parameters

Site	Membrane	Pressure (psi)	DOC (mg/L as C)	Color (cpu)	TH (mg/L as $CaCO_3$)
AID	U90-G50	100	11.92	23	309
AID	"	75	11.83	23	307
AID	"	50	12.53	22	310
AID	"	35	10.81	34	127
AID	"	50-reject	22.87	70	346
AID	OSMO PT-2	75	16.40	33	327
AID	"	50	14.19	33	329
AID	"	40	15.43	32	322
AID	"	30	13.76	33	329
AID	"	50-reject	14.75	35	322
AID	OSMO PT-4	50	14.21	32	331
AID	"	40	14.53	33	325
AID	"	30	14.65	33	321
AID	"	25	13.49	34	327
AID	"	50-reject	14.91	35	334
AID	UFP4040	100	13.25	35	330
AID	"	75	14.01	35	325
AID	"	50	12.60	32	334
AID	"	30	13.33	33	323
AID	"	50-reject	14.51	35	328
Raw			14.95	35	332

Table 97. Secondary Membrane Testing - THM Formation Curve Data (all values in μg/L)

Site	Membrane	Initial Cl_2 Dose ($mg/L\ Cl_2$)	Time (hours)				
			1	24	48	72	96
AID	U90-650	10	115	304	300	368	382
AID	U90-650	20	113	418	514	551	605
AID	Osmo PT-2	10	221	369	381	*	*
AID	Osmo PT-2	20	248	738	937	990	924
AID	Osmo PT-4	10	221	351	369	*	*
AID	Osmo PT-4	20	280	649	785	860	942
AID	UPF 4040	10	104	304	291	*	*
AID	UPF 4040	20	131	501	636	755	780

*Chlorine Residual has been 0.0 mg/L for a period greater than 24 hours.

Table 98. Secondary Membrane Testing - Chlorine Dissipation Curve Data

(all values in mg/L - Cl_2)

Site	Membrane	Initial Cl_2 Dose (mg/L Cl_2)	Time (hours)				
			1	24	48	72	96
AID	U90-650	10	3.5	0.18	0	--	--
AID	U90-650	20	13.1	7.9	5.9	5.5	4.7
AID	Osmo PT-2	10	2.4	0	--	--	--
AID	Osmo PT-2	20	11.6	7.4	5.1	3.5	3.0
AID	Osmo PT-4	10	2.2	0	--	--	--
AID	Osmo PT-4	20	11.5	5.8	4.1	3.4	1.9
AID	UPF 4040	10	3.9	0	--	--	--
AID	UPF 4040	20	13.6	7.3	5.7	4.4	3.0

Table 99. Parameter Removal and Retention Based on Molecular Weight

Membrane	MW	DOC		TH		Color		THM's	
	Passage	% Removal from raw	% Remaining in product water	% Removal from raw	% Remaining in product water	% Removal from raw	% Remaining in product water	Value µg/L	% Removal from highest value
BW30	< 100	95.5	4.5	95.7	4.3	96.9	3.1	32^a	96.6
N50	< 400	90.4	9.6	69.6	30.4	97.1	2.9	39^a	95.9
V90-G10	< 2,000	40.4	59.6	13.2	86.8	60.0	40.0	326^a	65.4
UFP-4040	< 10,000	15.7	84.3	0.0	100	8.6	91.4	780^b	17.2
U90-G50	< 20,000	16.2	83.8	6.6	93.4	37.1	62.9	605^b	35.8
PT-2	< 20,000	5.1	94.9	0.9	99.1	5.7	94.3	929^b	1.4
PT-4	< 40,000	5.0	95.0	0.3	99.7	8.6	91.4	942^b	--

a - 10 mg/L Cl_2 - initial chlorine dose

b - 20 mg/L Cl_2 - initial chlorine dose

Table 100. AID Test Matrix True Color vs. Feed Pressure and Product Recovery Rate
Raw Water Average = 37 cpu

Feed Pressure (psi)	Product Recovery Rate %			
	50	60	75	90
80	1.5	1.5	2.5	1.5
90	2.0	1.0	1.0	1.0
100	1.0	1.5	1.5	1.5
110	1.0	1.0	1.0	1.0

Table 101. AID Test Matrix Trihalomethane Formation Potential vs. Feed Pressure and Product Recovery Rate
Raw Water Average (without chlorination) = 0 μg/L

Feed Pressure (psi)	Product Recovery Rate %			
	50	60	75	90
80	42	36	32	29
90	23	25	30	30
100	18	28	33	49
110	29	40	43	60

Table 102. AID Test Matrix Total Dissolved Solids vs. Feed Pressure and Product Recovery Rate
Raw Water Average = 487 mg/L

Feed Pressure (psi)	Product Recovery Rate %			
	50	60	75	90
80	247	268	292	315
90	222	261	317	328
100	249	260	301	300
110	221	256	282	289

Table 103. AID Test Matrix Total Hardness vs. Feed Pressure and Product Recovery Rate
Raw Water Average = 319 mg/L as $CaCO_3$

Feed Pressure (psi)	Product Recovery Rate %			
	50	60	75	90
80	153	174	189	194
90	154	173	187	194
100	152	169	187	187
110	144	168	187	202

Table 104. AID Test Matrix Total Organic Carbon vs.Feed Pressure and Product Recovery Rate
Raw Water Average = 16.2 mg/L as C

Feed Pressure (psi)	Product Recovery Rate %			
	50	60	75	90
80	1.9	2.3	2.3	2.4
90	1.8	2.2	2.3	2.5
100	2.1	2.4	2.3	2.5
110	2.2	2.2	2.7	2.7

with increased product recovery rate. However all observed values are well-below the MCL of 0.1-mg/L. The noted trends may be explained by the factors discussed in the VOG analysis.

The parameter removal percentages may be seen in Table A-26 and Figures 115, 116, 117, and 118. The expected trend of decreasing removal rate based on raw water with increasing product recovery rate is seen. Comparison of this AID data with the corresponding VOG date shows that for both sites, the removal percentages are similar.

Long-term testing at AID was initiated on March 15, 1985, and terminated on May 31, 1985, after an elapsed run time of 1018 hours. Table A-27 and Figure 119 give a summary of the pilot plant operating data. Several trends may be noted from the data. From the time of the initiation of the run to approximately the 680 hour mark a gradual increase in product flux was noted. Flux improved gradually from 13.6 gpd/ft² to approximately 14.2 gpd/ft² over the first 360 hours. At that point, an operational change to a feed pressure of 100 psi was instituted. An immediate jump in flux to 14.9 gpd/ft² was noted followed by a gradual increase to approximately 15.5 gpd/ft². This condition persisted until a chemical cleaning was done at the 720 hour elapsed time. Subsequent to cleaning a dramatic flux increase to 20.8 gpd/ft² was accomplished. This flux declined gradually over the last 300 hours of operation to 20.0 gpd/ft².

The initial rise in flux prior to the cleaning is an interesting phenomenon with several possible explanations. The raw and feed water at AID generally has a lower DOC and true color value that at VOG. If the organic substances were mainly responsible for the blinding of the surface, a lower bulk solution would eventually effect a release of substances from the membrane surface and thus improve flux. Another explanation could be the gradual increase in water temperature seen as ambient temperatures increased. A third reason could be an actual feed pressure higher than the gauge pressure. Also the longer daily operation could have provided more membrane flushing and less time for bacteria to become attached during stagnant conditions.. It is difficult to ascribe any one reason to this slight but definite rise in permeate flux.

The dramatic increase in flux after cleaning shows that the fouling imposed by the VOG and AID feed waters was reversible. While the post-cleaning flux is greater than the original flux, a higher feed pressure was utilized after the cleaning (100 psi vs. 90 psi). Subsequent to cleaning a steady decline in flux to the 20.0 gpd/ft² level was experienced. Such a decline is to be expected given the new fouling of the membrane.

At AID the feed pressure did not vary substantially with time but minor fluctuations in the readings were found. This could be due to individual operator reading of the gauge. A similar caution should also be given to the reading of the flow gauges (which determine flux and percent product recovery). The gauges are manually controlled and read and different operators could read them differently. A difference in 0.1 gpm on the product flow gauge would mean a difference in 0.16 gpd/ft² in the flux value. These differences however are slight and unavoidable when

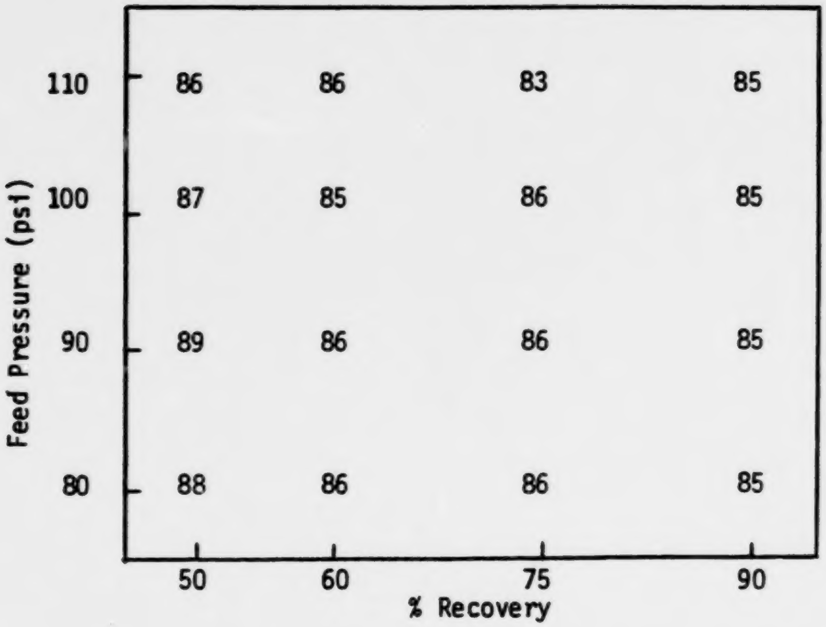


Figure 115 AID Test Matrix - Isopleths for percent DOC removal at varying feed pressure and product recovery.

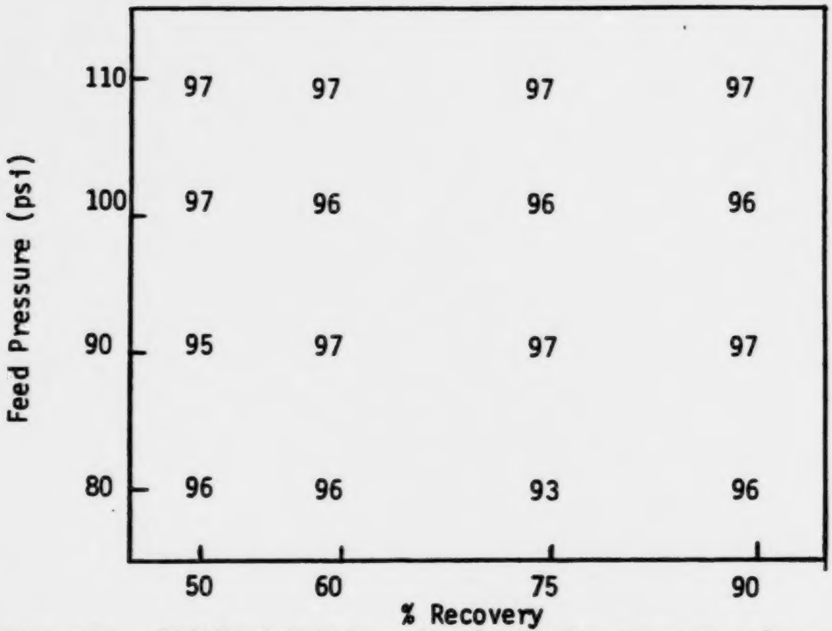


Figure 116 AID Test Matrix - Isopleths for percent color removal at varying feed pressure and product recovery.

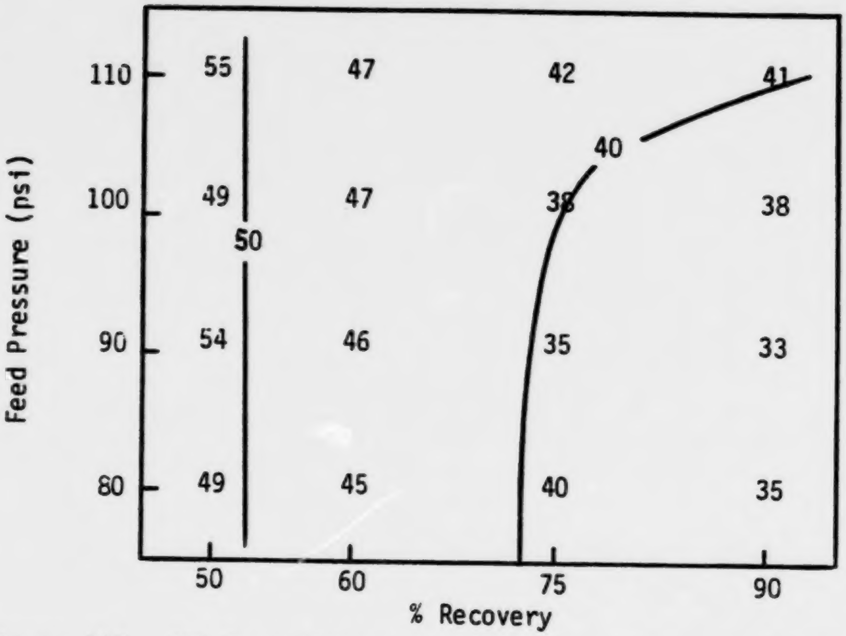


Figure 117 AID Test Matrix Isopleth for percent TDS removal at varying feed pressure and product recovery.

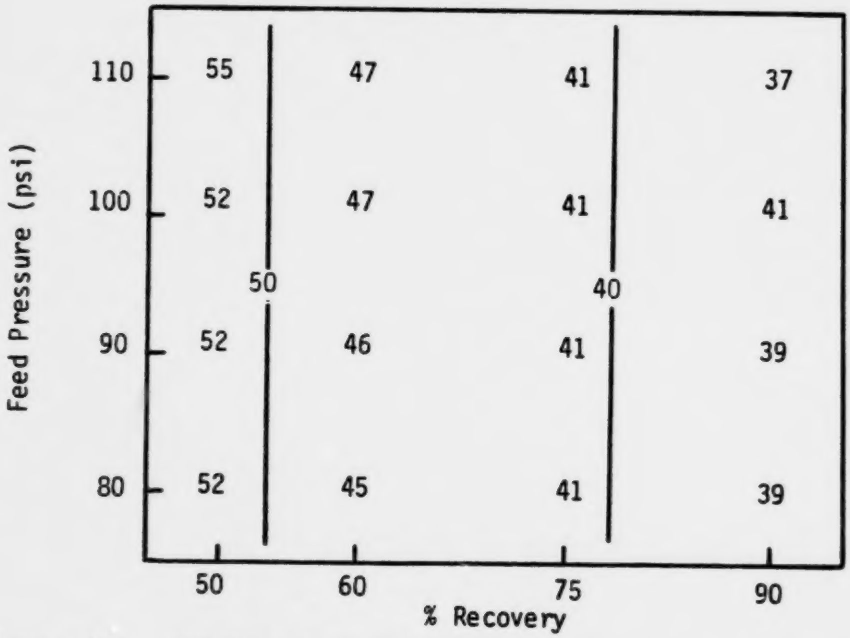


Figure 118 AID Test Matrix Isopleth for percent total hardness removal at varying feed pressure and product recovery.

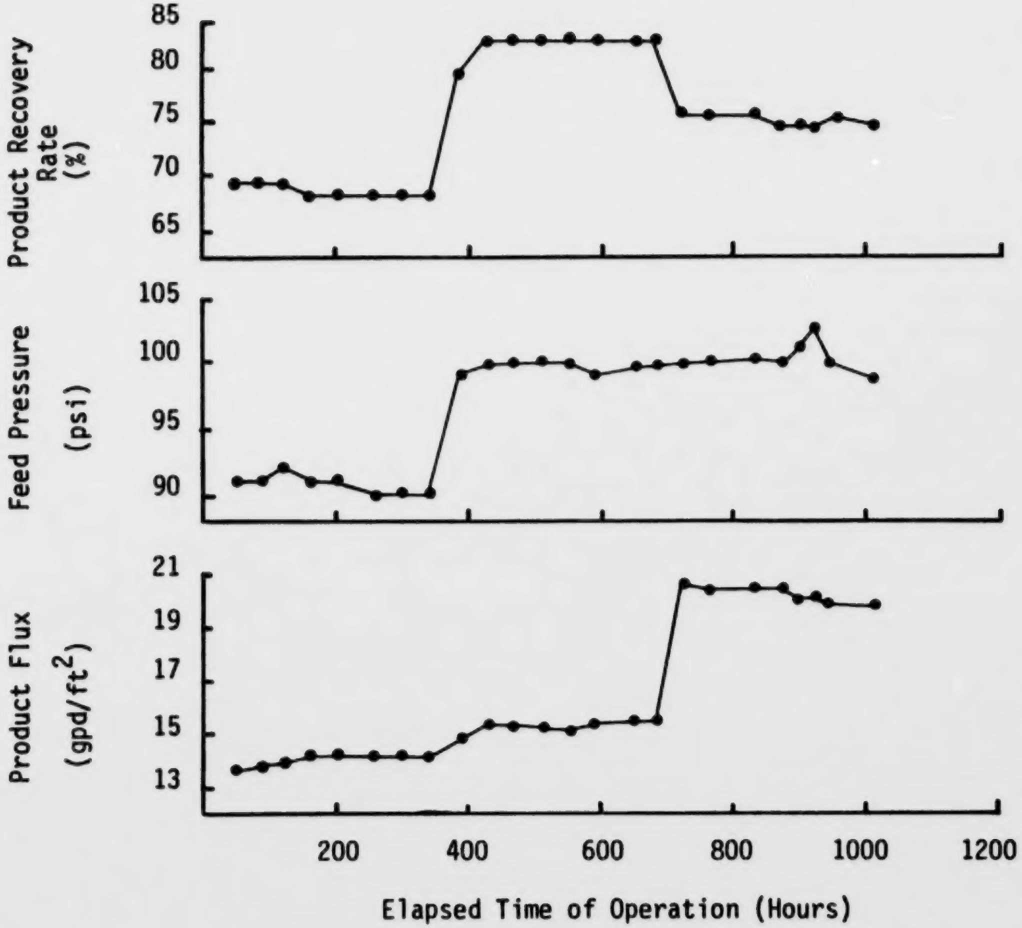


Figure 119 AID Pilot Plant Operation-Product Recovery Rate, Feed Pressure and Product Flux versus Time of Operation.

conducting a pilot plant operation on a continuous basis. Product recovery rates also did not vary substantially once a particular value was achieved. In most cases the product recovery rate was set by the limitations of the reject flow valves particularly when the reject flows were relatively low (insensitivity in valves).

To aid the discussion of various water quality parameters, several common tables have been provided. Tables A-28, A-29, A-30, and A-31 provided a data summary of the raw, feed reject and product water streams respectively. Table A-32 gives the TOXFP and THM's data. Tables A-33 and A-34 provide the percent parameter removals in the product water based on feed and raw water values, respectively. Tables 105 gives the correlation coefficients between the product water parameters values and the raw and feed values. Table 106 gives the correlation coefficients between the product water parameters.

The values for the raw, feed, and product water DOC values as a function of time are given in Figure 120. As can be seen the product water values are relatively low with a range of 1.2-mg/L to 2.9-mg/L. The average value of 2.2-mg/L is slightly higher then seen at VOG. Conversely, the average feed value at AID (29.3-mg/L) is slightly less. It is possible that a higher proportion of the organic carbon is present as low molecular weight substances. This would allow a greater passage to the product water.

Tables A-33 and A-34 and Figure 121 show that DOC is substantially removed in the product water as evidenced by the average 92% and 85% removal based on feed and raw water values, respectively. An interesting trend is noted in that removal percentages improved after the membrane cleaning. It would appear that the additional flux of water served to lower the product water quality through dilution.

Examination of Table 105 shows that a negative correlation exists between the product water DOC and the feed water DOC. Again, as in the case at VOG, the organic carbon is being removed via size restriction to low, relatively uniform values. It is also apparent the product water DOC is not well correlated with any other product water parameter.

The true color values for the raw, feed and product water streams are plotted on a time of operation scale in Figure 122. The observed product water value was fairly low with an average of 1.5 cpu. This average is slightly less than found at VOG. The average feed color is 60 cpu with a range of 45 cpu to 85 cpu. This feed value is much less than seen at VOG.

Examination of Tables 105 and 106 shows that product water color value not to be substantially linearly correlated with the feed water color or any other product water parameter. Once again the rather uniform low value and inherent analytical variation in reading these low colors would tend to obscure any trends. A similiar situation was also observed in the VOG data.

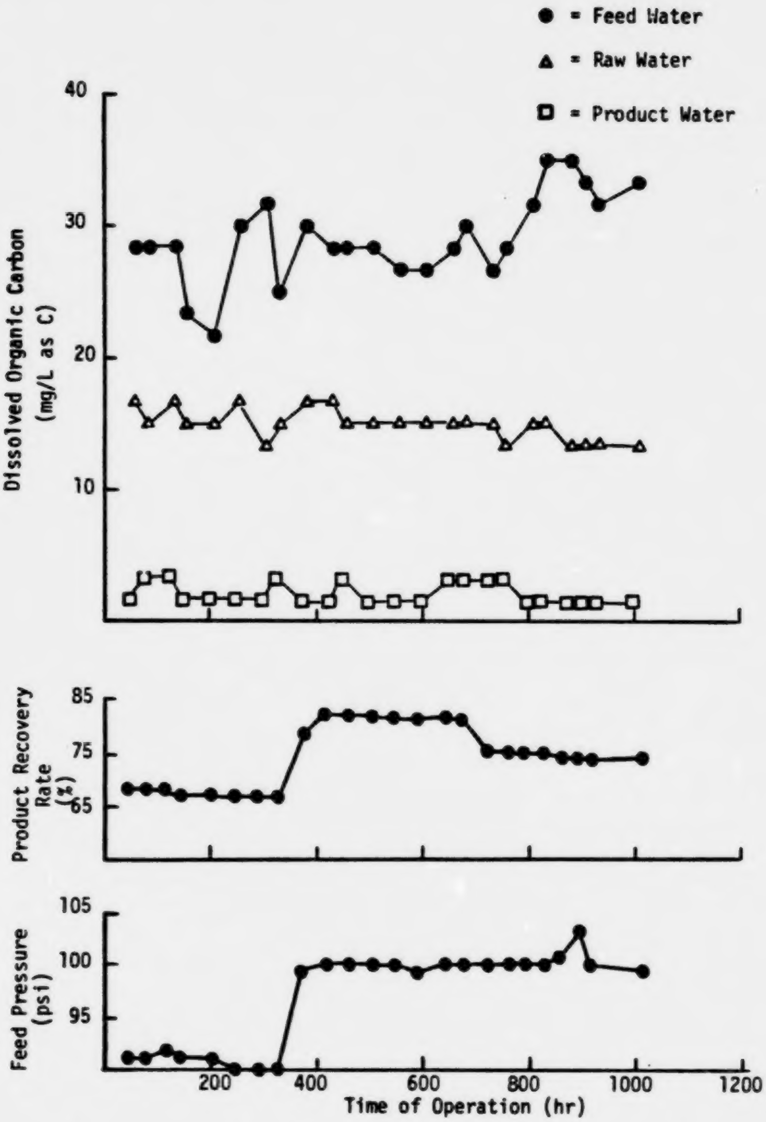


Figure 120 AID Membrane Pilot Plant Operation - Dissolved Organic Carbon versus time of operation, product recovery rate and feed pressure.

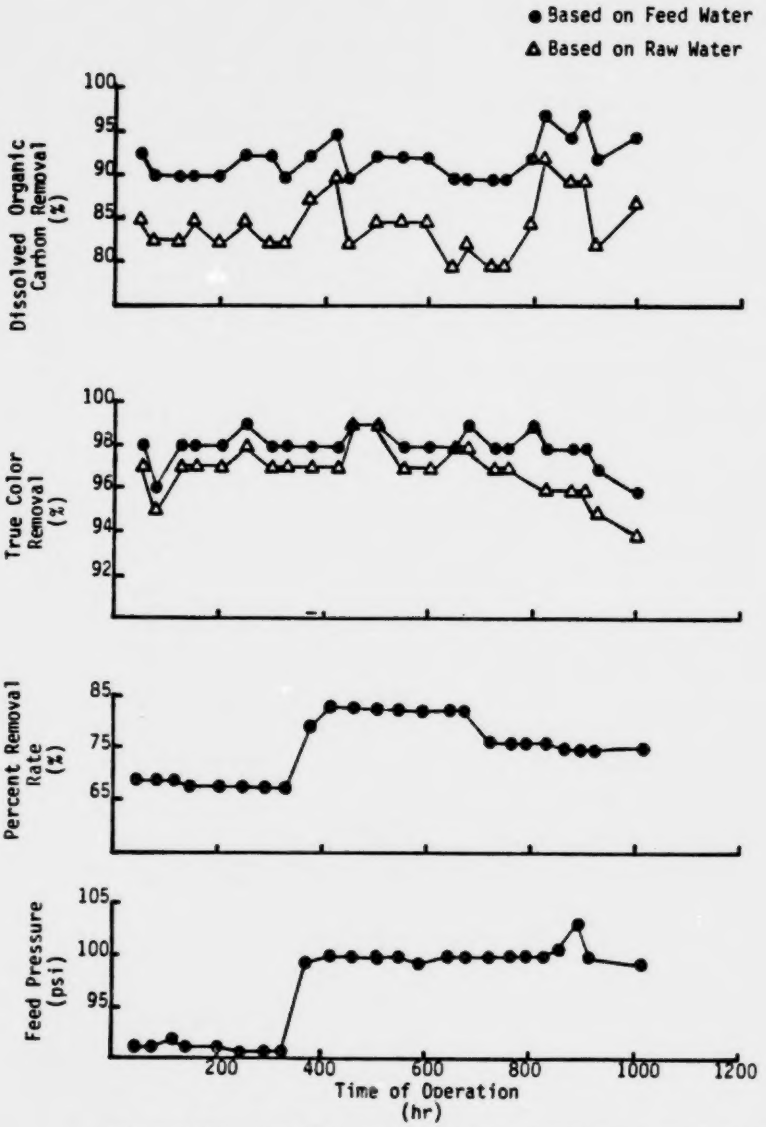


Figure 121 AID Membrane Pilot Plant Operation - Removal of Dissolved organic carbon and true color versus time of operation, product recovery rate and feed pressure.

TABLE 105. AID Linear Correlation Coefficients Product Water Parameters vs Feed and Raw Water Parameters

Product Water	Feed Water					Raw Water				
	DOC	Color	Alkalinity	T. Hardness	TDS	DOC	Color	Alkalinity	T. Hardness	TDS
DOC	-0.57	-0.29	-0.47	-0.40	0.10	0.17	0.2	0.18	0.04	0.24
Color	0.24	0.05	0.23	0.29	0.13	-0.19	-0.16	0.01	0.03	-0.31
Alkalinity	0.40	0.56	0.66	0.63	0.59	-0.16	0.13	0.40	-0.06	0.31
T. Hardness	0.56	0.29	0.71	0.86	0/61	-0.41	-0.20	0.15	0.35	0.31
TDS	0.19	0.19	0.43	0.31	0.53	-0.04	0.25	0.34	-0.11	-0.66
THMFP	0.37	0.28	0.58	0.71	0.76	-0.35	-0.01	0.34	0.15	0.65
TOXFP	0.55	0.21	0.81	0.78	0.60	-0.31	-0.24	0.02	0.20	0.37

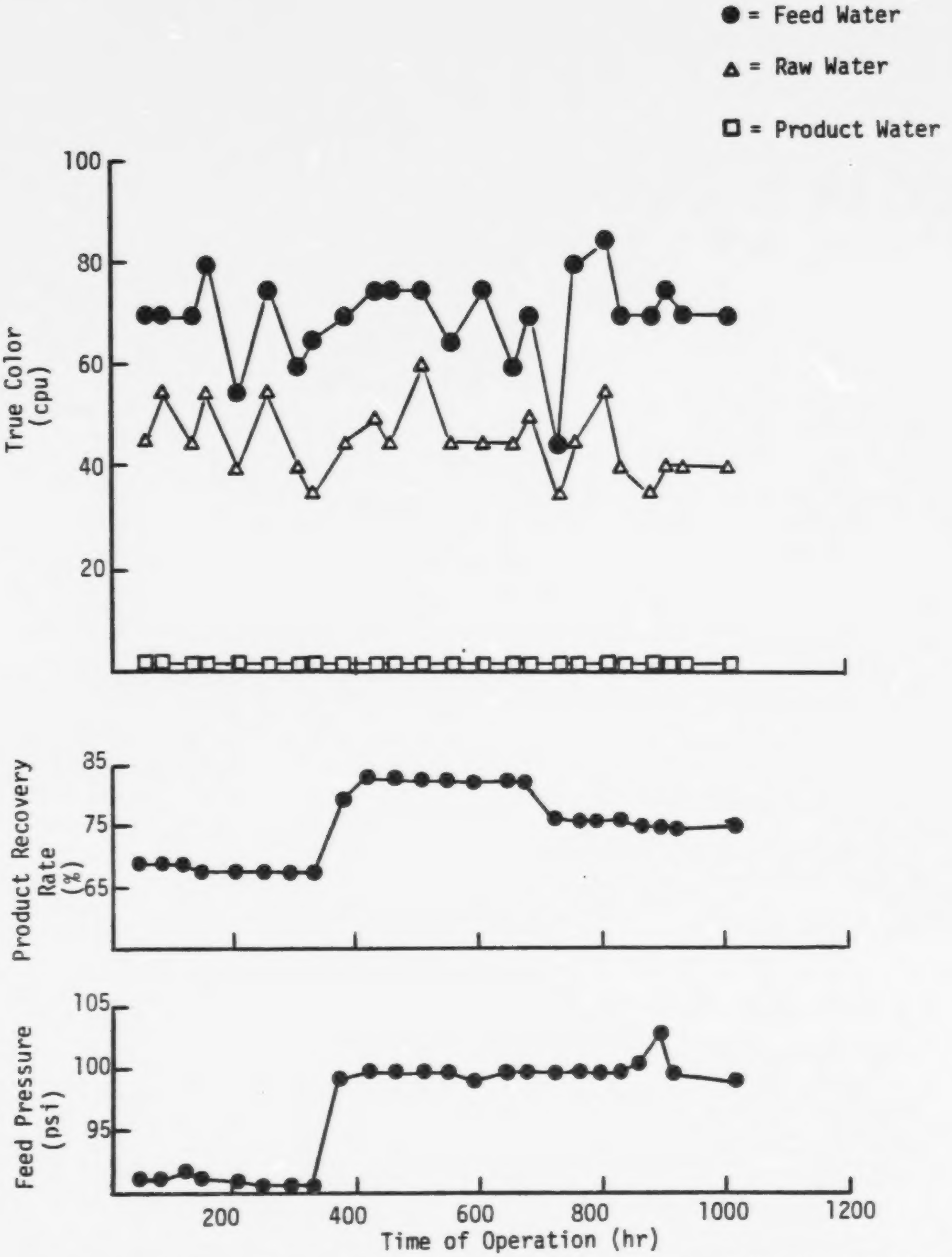


Figure 122 AID Membrane Pilot Plant Operation - True Color versus time of operation, product recovery rate and feed pressure.

TABLE 106. AID Correlation Coefficients - Product Water Parameters

Parameter	DOC	Color	Alkalinity	T. Hardness	TDS	THMFP	TOXFP
DOC	1.00	-0.33	-0.37	-0.37	0.07	0.04	-0.31
Color	-0.32	1.00	0.38	0.57	0.01	0.18	0.22
Alkalinity	-0.37	0.38	1.00	0.65	0.31	0.62	0.54
T. Hardness	-0.37	0.57	0.65	1.00	0.25	74	0.70
TDS	0.07	0.01	0.31	0.25	1.00	0.52	0.40
THMFP	0.04	0.18	0.62	0.74	0.52	1.00	0.70
TOXFP	-0.31	0.22	0.54	0.70	0.40	0.70	1.00

The THM's data are summarized in Table A-32 and Figure 123. As can be seen values are available for chlorine dosages of 7-mg/L and 10-mg/L. After several instances where no residual was detected at the end of 96 hours for the initial dose of 7-mg/L chlorine, the project personnel decided to use a 10-mg/L initial chlorine dose. In this manner a residual would always be expected and the THM's values obtained at higher dosages could be observed. In many cases a residual of 3.0-mg/L or greater was found. Later in the study, a comparison sample chlorinated at 7.0-mg/L was done for several of the sampling dates. In these cases, a residual greater than 0.2-mg/L was always found.

In all cases except one the THM's values were comfortably below the current MCL of 0.10-mg/L. In the one exception the recorded value at an initial 10-mg/L Cl_2 dose was 113 µg/L, slightly over the MCL. This value occurred about 3 hours after the chemical cleaning. At the same time the highest DOC value was noted. Since the blinding of the membrane could help in filtration, immediately after a cleaning a greater solute flux could occur until the membrane blinds again. Also residual contamination is a possibility. In any case, for future operations, it might be necessary to blend the water with some pre-wash product water to ensure compliance with the MCL.

Examination of Tables 105 and 106 shows that THM's is not substantially linearly correlated with any feed water organic parameter or with any other product water parameter except TOXFP. It is believed that the inherent variations present in the chlorination, incubation and analytical methods tend to obscure any trends, especially at relatively tow THM's values.

The TOXFP data for laboratory chlorinated water may be seen in Table A-32 and Figure 124. While there is no MCL for TOXFP all samples were below the 100 µg/L as Cl level. In many cases the recorded value was between 30 and 60 µg/L as Cl. Examination of Tables A-32 and 140 show that TOXFP is substantially correlated to only the product water THM's value as would be reasonably expected.

The TOXFP:THM's ratio has an average of 0.95. This value is lower than the value found at VOG. It is interesting to note that at VOG the predominant THM's species was chloroform while at AID the dichlorobromo- and dibromochloro-forms of methane tended to dominate the THM speciation. Although the THM speciation was different at VOG and AID, the THMFP removal was similar.

The data for the raw, feed and product water TDS values are plotted in Figure 125. The product water value has an average of 282-mg/L while the feed and raw TDS average values are 645-mg/L and 491-mg/L respectively. These value are greater than the values observed at VOG. However, the product water value is comfortably below the recommended limit. It is also interesting to note that the product water value exhibited a relatively wide range. The average raw and feed water removals were 56% and 43% (See Figure 126).

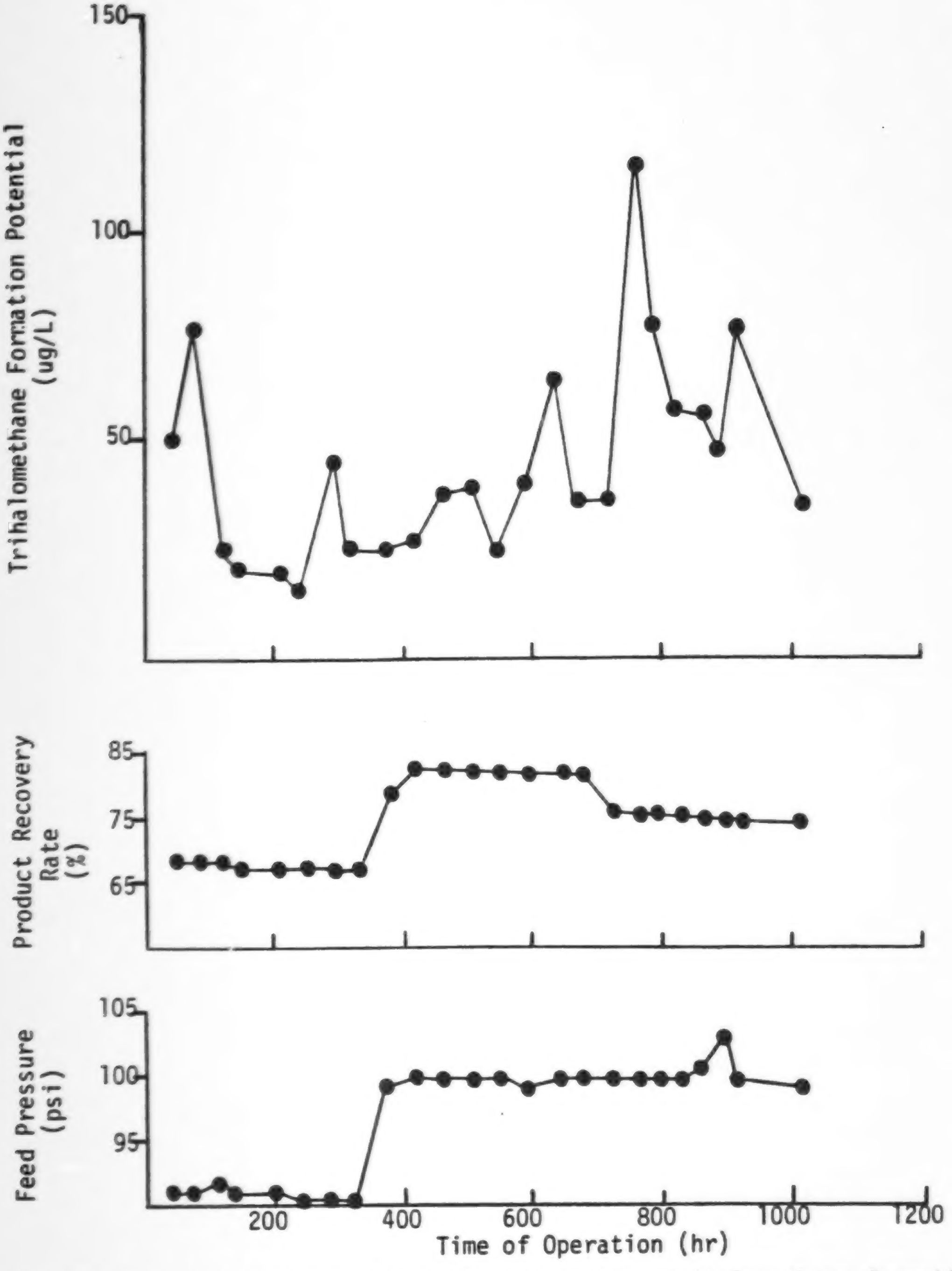


Figure 123 AID Membrane Pilot Plant Operation Trihalomethane formation potential versus time of operation, product recovery rate and feed pressure.

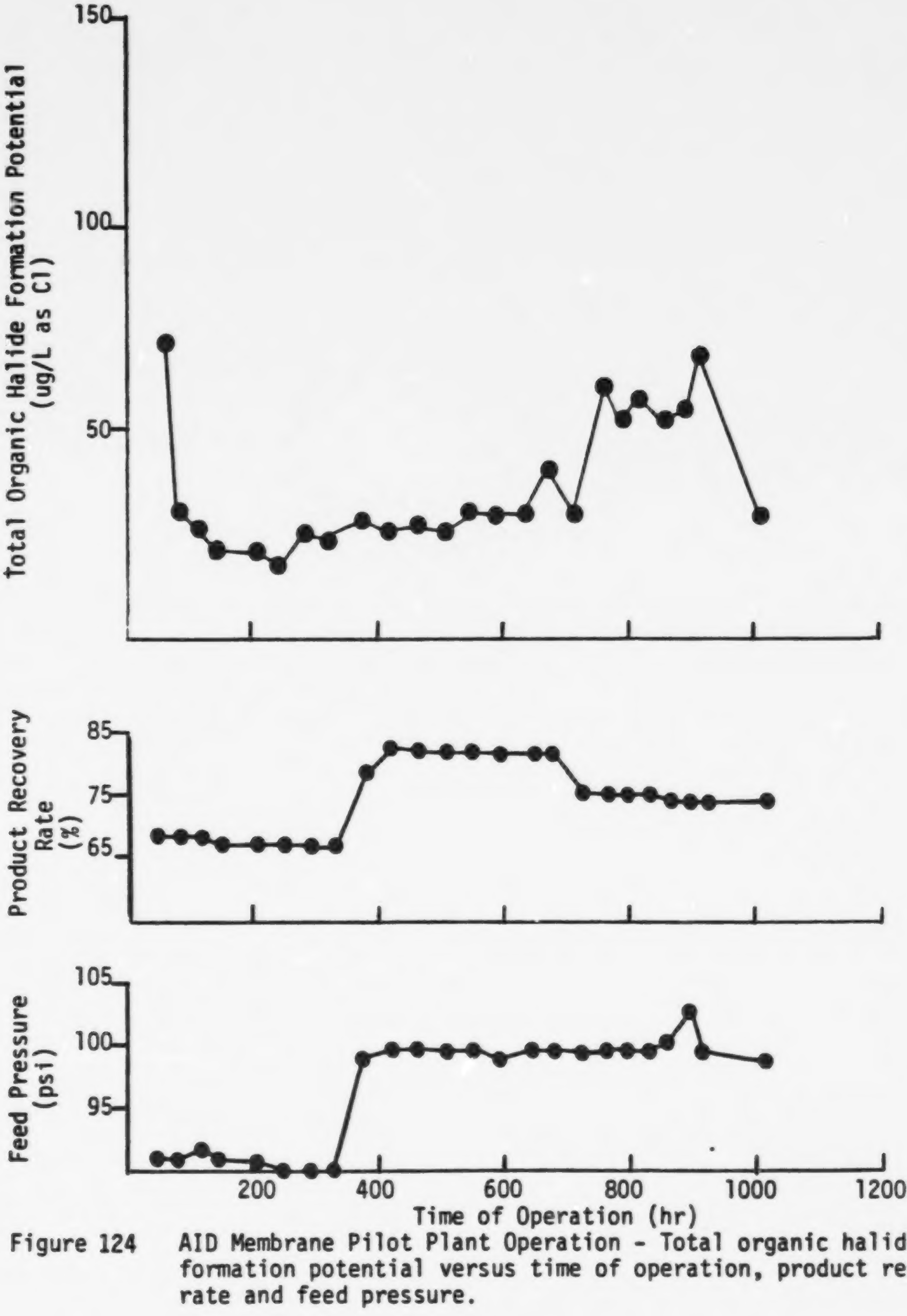


Figure 124 AID Membrane Pilot Plant Operation - Total organic halide formation potential versus time of operation, product recovery rate and feed pressure.

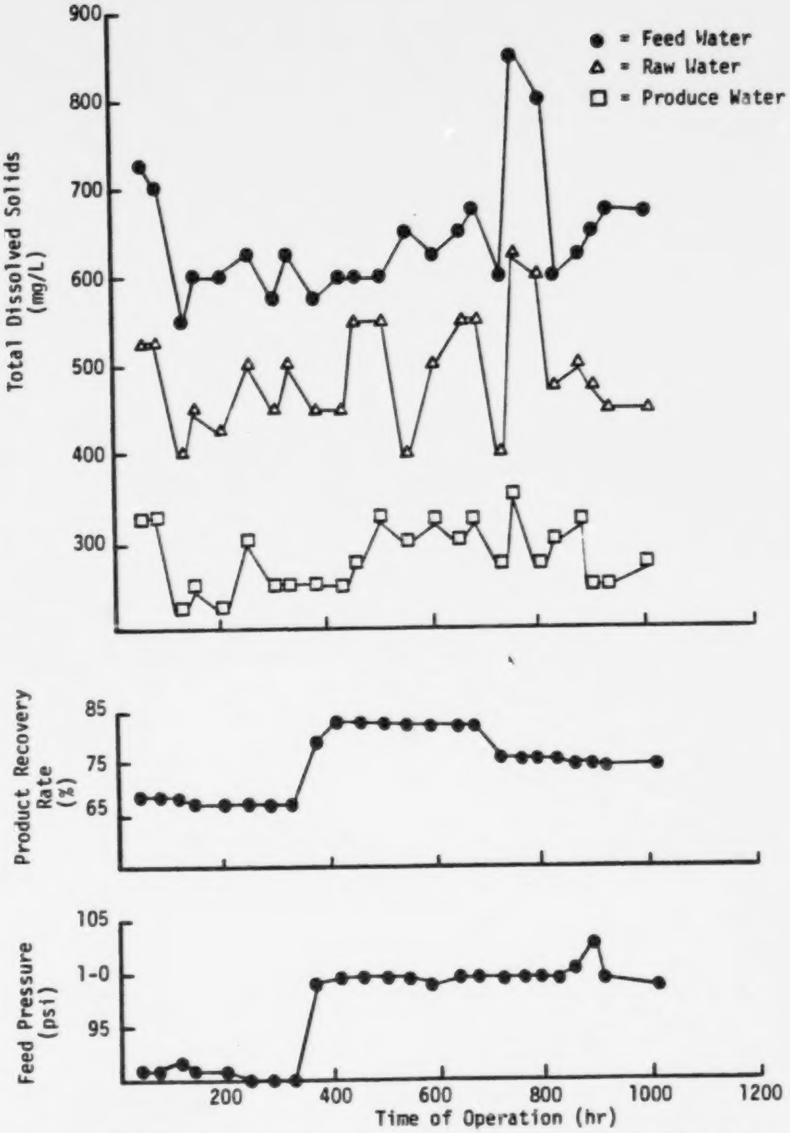


Figure 125 AID Membrane Pilot Plant Operation - Total dissolved solids versus time of operation, product recovery rate and feed pressure.

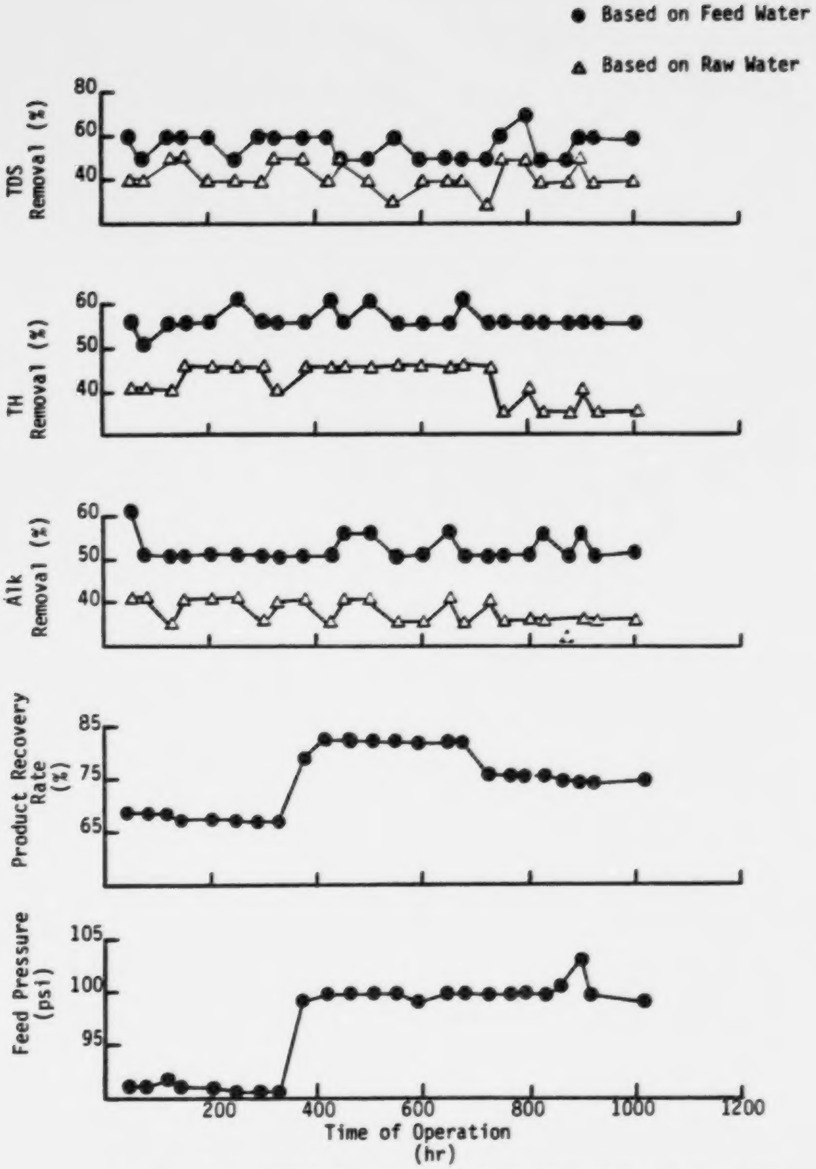


Figure 126 AID Membrane Pilot Plant Operation - Removal of total dissolved solids, total hardness and alkalinity versus time of operation, product recovery rate and feed pressure.

Examination of Tables 105 and 106 shows only a moderate correlation with the feed and raw TDS value. This situation differs from the situation seen at VOG where the correlation was stronger. It would appear that the situation with correlations would be site-specific for a given source water.

The TH data plots for the raw, feed, and product streams may be seen in Figure 127. The product water average value is approximately 187-mg/L as $CaCO_3$ with a relatively narrow range of 172-mg/L as $CaCO_3$ to 207-mg/L as $CaCO_3$. The raw water values have a relatively narrow range of 304-mg/L as $CaCO_3$ to 329-mg/L as $CaCO_3$ and reflect the limestone layers that serve as the majority supply source. The product water value is approximately twice the value at the current treatment plant. While the water would be considered "hard" it is certainly potable. The product water exhibit removals averaging 56% and 40% based on feed and raw water respectively (see Figure 126).

From Table 105 it can be seen that the product water TH value is substantially correlated with the feed water value. This situation did not occur to the same degree at VOG. At AID the decreased organic concentration and higher TH could lessen the chance of chemical complexation and chelation. The quality of the product water level could then vary more directly with the quality of the feed.

The data plots for the raw, feed and product alkalinity values may be seen in Figure 128. The product water average was 180-mg/L as $CaCO_3$ while the feed and raw water averages were 372-mg/L as $CaCO_3$ and 286-mg/L as $CaCO_3$ respectively. The alkalinity in the product water provided an adequate buffering capacity so that the final pH after aeration and chlorination was in the pH 7.0 to pH 7.5 range. Typical percent removals from the feed water values were in the 49% to 55% range. (See Figure 126) Also, from Tables 105 and 106, the product water alkalinity is moderately correlated to the feed water alkalinity and to the product water TH.

During the testing period at AID samples were periodically taken from the individual pressure vessels each of which held three membranes. The data are summarized in Tables 107, 108, and 109. Tables 107 and 108 provide the individual parameter data while Table 109 provides averages of the parameters for the individual vessels and certain combinations. As earlier stated the incoming feed flow is split between 1A and 2A whose reject waters become the feed water for 1B and 2B respectively. During the test period, it was determined that the flows in the 1A-1B series were similar in flow rate and water quality to the flows in the 2A-2B series.

Examination of the data in Table 109 shows that the water quality from the 2B pressure vessel to be substantially better than the other series. It would also appear that the water quality in 2A is slightly better than in either 1A or 1B. The dramatic improvement in 2B is particularly evident in the inorganic parameters. Given that the feed water quality parameter levels to the "B" series would be slightly higher one might expect the product water quality to be slightly less. The 1B vessel shows this to a certain degree, but not the 2B vessel.

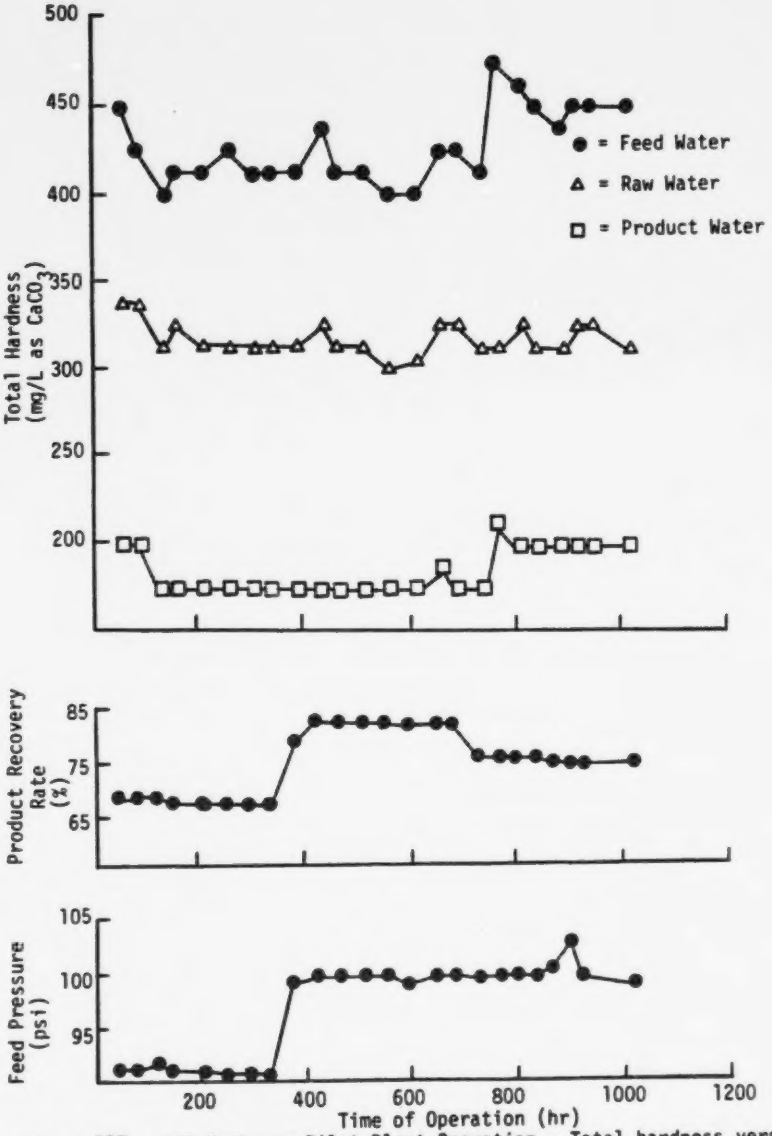


Figure 127 AID Membrane Pilot Plant Operation - Total hardness versus time of operation, product recovery rate and feed pressure.

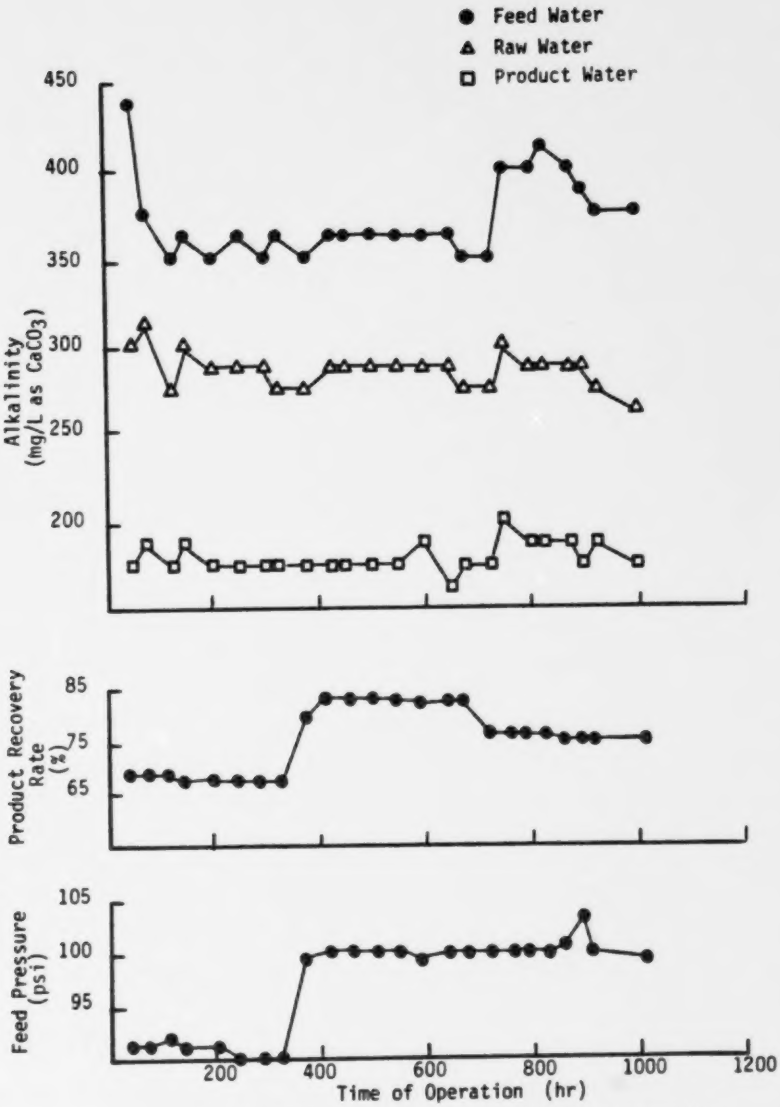


Figure 128 AID Membrane Pilot Plant Operation. Alkalinity versus time of operation, product recovery rate and feed pressure.

Table 107. AID Individual Pressure Vessal Product Water Quality Data Summary (continued)

Values reported as $\frac{\text{1A 1B}}{\text{2A 2B}}$

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	% Product Water Recovery (%)	DOC (mg/L as C)		Color (cpu)		Alkalinity (mg/L as $CaCO_3$)		Total Hardness (mg/L as $CaCO_3$)		Total Dissolved Solids (mg/L)	
3-20	81	91	69	2.5	1.8	1.0	2.5	199	216	225	221	332	315
				2.3	2.8	4.0	2.5	190	143	206	149	269	214
3-26	149	91	68	2.5	2.7	1.5	2.0	202	219	202	160	277	312
				2.3	2.3	1.0	1.0	163	133	206	125	253	196
4-1	249	90	68	2.5	2.4	1.0	0.5	191	206	198	177	294	342
				1.4	1.5	1.0	0.5	166	134	163	159	234	208
4-7	332	90	68	2.1	3.0	1.5	1.0	206	197	200	204	234	274
				2.1	1.9	1.0	0.5	172	122	164	117	192	204
4-14	417	100	83	1.7	1.6	2.0	1.5	185	196	211	202	200	278
				2.0	1.4	1.5	1.0	159	115	168	113	242	204

(continued)

Table 107 (continued). AID Individual Pressure Vessal Product Water Quality Data Summary

Values reported as $\frac{1A\ 1B}{2A\ 2B}$

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	% Product Water Recovery (%)	DOC (mg/L as C)		Color (cpu)		Alkalinity (mg/L as $CaCO_3$)		Total Hardness (mg/L as $CaCO_3$)		Total Dissolved Solids (mg/L)	
4-20	505	100	83	2.6	2.9	0.5	0.5	197	193	198	204	351	345
				2.1	2.0	1.0	0.5	155	115	162	111	289	236
4-26	591	99	83	2.4	2.4	1.5	2.0	197	201	191	192	284	305
				2.2	1.8	1.5	0.5	157	117	157	112	243	157
5-3	676	100	76	2.6	3.0	1.0	1.0	189	197	202	194	300	308
				2.4	2.9	1.0	1.0	155	113	166	117	279	201
5-9	759	100	76	2.5	2.9	1.0	1.5	243	231	261	249	468	473
				2.4	1.6	1.0	1.0	206	118	221	129	421	303
5-15	830	100	75	1.4	1.6	2.5	2.5	230	217	239	232	282	262
				1.0	0.8	2.5	1.5	192	113	202	116	281	191
5-21	895	103		1.3	1.4	2.5	2.0	236	209	243	226	342	258
				1.1	1.1	2.0	1.0	180	113	195	153	214	129

Table 108. AID Individual Pressure Vessel Product Water THM's and TOXFP Data Summary Values From Each Pressure Vessel Are Organized $[\frac{1A/1B}{2A/2B}]$. THM's in µg/L and TOXFP in µg/L as Cl

Sample Date	THM's Chlorine Dose 7 mg/L Cl_2	 10 mg/L Cl_2	TOXFP Chlorine Dose 7 mg/L Cl_2	 10 mg/L Cl_2
3-20	NA	29/43 23/24	NA	62/76 68/53
3-26	27/10 14/18	NA	25/21 31/10	NA
4-7	NA	22/19 26/20	NA	19/21 17/19
4-14	NA	40/28 24/41	NA	32/37 31/28
4-20	NA	27/33 29/24	NA	36/38 32/24
4-26	NA	57/49 38/30	NA	36/43 39/28
5-3	NA	39/32 37/39	NA	35/43 36/34
5-9	81/73 87/68	NA	NA	NA
5-15	NA	40/43 25/37	NA	46/40 33/33
5-21	NA	37/39 41/39	NA	44/54 43/35

Table 109. AID Individual Pressure Vessel - Product Water Parameter Averages[1]

Sample	DOC (mg/L C)	Color (cpu)	ALK (mg/L as $CaCO_3$)	TH (mg/L as $CaCO_3$)	TDS (mg/L)	THM's[2] (µg/L)	TOXFP[2] (µg/L as Cl)
1A	2.2	1.5	207	215	307	36	39
1B	2.3	1.5	207	206	316	36	44
2A	1.9	1.5	172	183	265	30	37
2B	1.8	1.0	121	127	204	32	32
1A + 1B	2.3	1.5	207	211	312	36	42
2A + 2B	1.9	1.3	147	155	235	31	35
1A + 2A	2.1	1.5	190	199	386	33	38
1B + 2B	2.1	1.3	164	167	260	34	38

[1] $\frac{1A/1B}{2A/2B}$

[2] 10 mg/L Cl_2 dosed samples only.

Several possible answers to this water quality difference were explored. Feed water quality to the "B" vessels was examined and found to be very similar. Feed pressures and transmembranic pressures drops were measured and found to be relatively similar. These observations tended to eliminate the possibility of a dramatic hydraulic or pressure related operational difference. After consultation with the membrane manufacturer it was decided that the 2B vessel contained a membrane(s) that had a more restricted molecular weight cutoff. Since the manufacture of the membranes has inherent variations, this appears to be the most plausible explanation.

In conjunction with the water quality parameters discussed in the previous section, the pressure drop across the individual pressure vessals should be examined. Subsequent to the membrane cleaning on May 6, 1985 the system was altered so that pressure readings could be taken for each individual pressure vessel product water. The data may be seen in Table 110. As can be seen the feed pressures for both "A" vessels and "B" vessels are not equal. The lower feed pressure for both "B" vessels is due to the feed-pressure loss in travel through the feed and brine channels of the "A" membranes.

Of particular interest is the comparisons of pressure drops between the "A" and "B" vessels and the difference between the #1 and #2 sets of pressure vessals. As can be seen the pressure loss through the "B" vessels is greater than the loss throught the "A" vessels. The difference is generally in the 8 psi to 12 psi range (or 3 psi to 4 psi per membrane element). The pressure difference can be generally ascribed to the difference in feed concentrations and the resultant differences in the concentration polarization layer. In the "B" vessels the feed concentration has been elevated due to the removal of relatively less concentrated product water. Such an elevated concentration would increase the concentration polarization layer and also serve to "blind" the membrane quicker.

There are also some differences in the noted pressure drops between the #1 and #2 series. In the "A" series the pressure drop in the #1 series is generally 1 psi to 2 psi greater. In the "B" series the "2" series exhibits a larger pressure drop usually in the 2 psi to 3 psi range. Such a situation would be consistent with the hypothesis that the #2-B pressure vessel contains membrane(s) that are "tighter" than the membranes located in the other vessels. Also slight differences could be due to the relative heights of the pressure vessels. Finally it should be noted that for all pressure vessels a slight increase in the pressure drop is noted with time. Such a phenomenon is to be expected with the graudal blinding of the membrane.

6.4.8.4 Summary

In summary many of the statements concerning the testing at VOG can be made for AID. In almost all cases the THM's MCL was easily meet even with a substantial chlorine residual. The product water had a low color and DOC content that would not require chlorine for bleaching of color or the elimination of taste and odor problems. The water is "harder" than the present product water at the site but is well within desirable potable

Table 110. AID Membrane Pilot Plant Test - Pressure Values for the Individual Pressure Vessels (all values in psi)

Sample Date	Feed Pressure	1A	2A	ΔP		Feed Pressure	1B	2B	ΔP	
				1A	2A				2B	2B
5-09	100	69	70	31	30	90	53	50	37	40
5-10	100	68	69	32	31	90	50	49	40	41
5-11	100	67	68	33	32	91	51	50	39	40
5-12	100	68	69	32	31	91	52	50	39	41
5-13	100	68	69	32	31	91	52	50	39	41
5-14	99	68	70	31	29	90	52	51	38	39
5-15	100	68	70	32	30	91	52	51	39	40
5-16	101	68	70	33	31	91	52	51	39	40
5-17	101	68	69	33	32	92	51	50	41	42
5-18	101	68	70	33	31	92	51	49	41	43
5-19	101	69	69	32	32	92	51	49	41	43
5-20	102	68	69	34	33	93	52	50	41	43
5-21	103	68	67	35	36	94	50	48	44	46
5-22	100	66	68	34	32	92	51	49	41	43
5-23	100	67	67	33	33	91	50	48	41	43
5-24	100	66	67	34	33	91	50	48	41	43
5-25	100	66	67	34	33	91	50	48	41	43
5-26	100	67	67	33	33	91	51	48	40	43
5-27	100	66	67	34	33	91	50	48	41	43
5-28	100	66	66	34	34	91	50	48	41	43
5-29	99	66	67	33	32	90	50	47	40	43
5-30	99	67	67	32	32	91	50	47	41	44
5-31	99	66	67	33	32	90	49	46	41	44

limits. It is also evident that exact predictions of water quality cannot be made from the feed water quality. At best it would appear that a range of values could be established for the product water quality.

A final point can be made concerning the cleaning of the membrane. From the cleaning accomplished at the AID site, it was found that the original flux could be restored. However, it was also found that product water quality will degrade slightly after cleaning. In a large-scale operation this effect could be minimized by not cleaning all membranes simultaneously.

6.4.9 Cost Data Estimates

At the current time, UF is not used as a unit process in the treatment of potable water to the author's knowledge. Thus, actual construction, operation and maintenance costs are unavailable and must be estimated for projected facilities. While pilot plant operation can provide data for the cost of operations, the differences in construction between a pilot plant and a permanent facility are too great to allow accurate scale-up factors.

Fortunately, a number of RO facilities are currently operating in the United States and other regions of the world. The physical set-up of the two processes would be quite similar as would be the labor required. Both processes require similar piping and valving, pressure vessels, chemical feeds, storage tanks, pumps and electrical instrumentation. Labor requirements would be similar due to similarity of the process in terms of operation. To this end the construction costs reported by (Gumerman et al., 1979) may be used with appropriate scaling for 1985 dollars.

For the construction costs to be used, it is assumed that spiral-wound UF membranes will be commercially available in the sizes available for RO membranes. It is further assumed the cost of the membrane will be compatible with the current price of RO membranes. From discussions with the manufacturer, such assumptions are reasonable. Furthermore it should be recognized that given the lower operating pressures, certain savings could be achieved in piping material requirements and pumps.

One main area of divergence in costs between an RO and UF installation would be the required electrical costs for pumps. The cost curves from Gumerman et al. (1979) assume an operating feed pressure of 400 psi to 450 psi with a water product recovery rates of 75% for 0.1 mgd to 1.0 mgd installations to 80% for facilities in the 1.0 mgd to 10 mgd range. The projected UF operation would operate at approximately 90 psi to 100 psi feed pressure with a product recovery rate to 75% to 80%. Thus the base power requirement estimate will be 25% of the previously given RO value.

In order to update the 1978 dollar values to 1985 dollars certain indexes are utilized. These include the Consumer Price and the Producer Prices Indices and the Engineering News Record Index. Table 111 gives the various cost categories, the applicable indices and comparison of the October 1978 data and the April 1985 data. As can be seen costs have risen approximately 50% since the October 1978 base period.

TABLE 111. Cost Estimate Categories and Applicable Index

Cost Equipment	Index	October 1978 Value	April 1985 Value	Factor 10/78 4/85
Manufactured Equipment	BLS General Purpose Machinery and Equipment - Code 114	221.3	320.5	1.448
Labor	ENR Skilled Labor Wage Index (1967 Base)	247	378	1.530
Electrical and Instrumentation	BLS Electrical Machinery and Equipment - Code 117	167.5	253.2	1.512
Housing	ENR Building Cost Index (1967 base)	255	359	1.408
Electrical Power	BLS Consumer Price Index Detailed Report	208.1	362.6	1.742

In Table 112, the estimate construction costs for a 0.1 MGD, 1.0 MGD and 10.0 MGD UF plant are given. As expected an economy of scale would exist for the larger-sized plants. The costs for the three plants are \$173,600, \$1,002,100, and \$7,190,300 for the 0.1 MGD, 1.0 MGD, and 10 MGD plant respectively.

Table 113 gives the estimated operation and maintenance costs for the three projected plants. It is important to note that prefiltration costs, chemical conditioning costs (acid and/or antiscalant), and ultimate disposal costs have not been calculated. These costs would be highly site-dependent. It should also be remembered that the great majority of the maintenance materials cost includes membrane replacement ever three years. From discussions with the manufacturer, it is expected that membrane life could be reasonably expected to be greater than three years. Thus, the materials costs will be somewhat conservative. The O&M costs would be \$19,400, \$171,600, and \$1,350,300 for the 0.1 MGD, 1.0 MGD, and 10 MGD plants, respectively.

The construction costs and O&M costs may also be calculated on a per 1000 gals basis. The construction costs were considered to be amortized over a 20-year period at an interest rate of 8%. The combined costs for the various plants were as follows:

0.1 MGD = \$0.48 + \$0.53 = \$1.01/1000 gals

1.0 MGD = \$0.29 + \$0.47 = \$0.76/1000 gals

10 MGD = \$0.20 + \$0.37 = \$0.57/1000 gals

It should be remembered that the need for prefiltration, chemical additions, and ultimate disposal would raise the cost while a longer than estimated membrane life would decrease the cost.

6.4.10 Summary.

Bench (1000 gpd) and pilot scale (25,000 gpd) investigations of membrane processes were conducted at two sites, VOG and AID, that utilized ground water supplies. Bench scale investigations were also conducted at one site, Olga, that utilizes a surface supply. Initially, one RO and six UF (low pressure) membranes were purchased and tested for product (permeate) water quality on a bench-scale level. The UF membranes are designed for operation up to 100 psi, whereas the RO membrane is intended to operate at 200-250 psi. Bench results from all three sites demonstrated the only membranes capable of producing a water from these highly organic sources that would meet the THM MCL and maintain a Cl_2 residual was the Filmtec UF membrane, (N-50) and the RO membrane (BW 3030).

The nominal molecular weights rejected by each membrane were supplied by each manufacturer and ranged from 40,000 to 100. Bench scale testing indicated that a molecular weight rejection of 2,000 would typically pass 50% of the raw DOC but only 20% of the color. The resulting product THMFP was generally 800 µg/L for the surface water source and 400 µg/L for either

TABLE 112. Estimate Construction Costs for Ultrafiltration Plant in April 1985 Dollars

Cost Category	Plant Capacity 0.1 MGD	 1.0 MGD	 10.0 MGD
Manufactured Equipment	$110,600	$ 647,300	$ 4,720,500
Labor	23,400	102,500	504,900
Electrical and Instrumentation	15,400	95,000	702,300
Housing	8,400	84,400	608,300
Subtotal	$157,800	$929,200	$6,536,600
Miscellaneous and Contingency[1]	15,800	92,900	653,700
Total	$173,600	$1,022,100	$7,190,300
Cost per 1000 gals[2]	$0.48	$0.29	$0.20

[1] 10% of Subtotal Value

[2] 20 year amortization at 8%

TABLE 113. Estimated Operations and Maintenance Costs for Ultrafiltration Plant in April 1985 Dollars

Cost Category	Plant Capacity 0.1 MGD	1.0 MGD	10 MGD
Energy Costs - Building	$ 800	$ 5,500	$ 43,900
Process	3,400	31,500	288,500
Maintenance Materials	13,200	131,800	1,013,600
Labor	2,000	2,800	4,300
Total	$19,400	$171,600	$1,350,300
Cost per 1000 gpd	$0.53	$0.47	$0.37

ground water source. The UF with molecular weight rejection of 400 passed less than 10% of the DOC and 3% of the color at any site, while typically producing a water with a THMFP of less than 50 µg/L at the ground water sites and approximately 100 µg/L for the surface water. The product was essentially colorless and the DOC was 2-mg/L or less at all three sites. The color, DOC, and THMFP of the RO membrane were approximately equal to the color, DOC, and THMFP of the best UF membrane. However, the RO membrane operated at 200 psi and rejected species of MW 100 or greater as opposed to the UF membrane which rejected species of MW 400 or greater at a pressure of 100 psi. The RO membrane rejected a much higher inorganic fraction than did the UF membrane. The RO membrane rejected over 90% of the TDS, total hardness (TH), Cl^-, and Na^+ at all sites, whereas the UF membrane inorganic rejection varied from 50% to 70% for the various parameters. Since the N-50 was the membrane that operated at the lowest pressure and still produced a water that met the THM MCL it was selected for extended operation.

The N-50 UF was installed at Olga in a bench scale unit capable of producing 1000 gpd and operated for 740 hours over a 45-day period. An operational percentage of recovery and feed pressure matrix was developed to determine the extended study operating conditions. Over conditions of 60 to 120 psi and 60% to 90% recovery water quality was best at high pressure (105+ psi) and lower recovery (60%). The matrix results indicated the THM MCL could be met if the operation conditions were 105 psi with 60% recovery. At these conditions product water quality and percent rejections were 1.6-mg/L DOC as C (92%), 3 cpu color (93%), 172-mg/L TDS (64%), 78-mg/L as $CaCO_3$ TH (65%), 60-mg/L Cl^- (40%), and 68-mg/L as $CaCO_3$ Alk (59%) with pH 7.8. During the extended operation, pressure was varied from 60 to 100 psi with percent recoveries of 60 to 90. The THM MCL was met for 105 psi, 60% recovery and for 75 psi, 60% recovery immediately after the membrane was chemically cleaned. The flux decreased with time during the extended study from 18 to 14 gpd/ft^2 over 150 hours of operation. After cleaning with the pressure at 95 psi, the flux declined from 22 to 16 gpd/ft^2 but the product quality remained constant. Flux was independent of recovery during the extended study. The water quality and percent rejection (%R) for the conditions meeting the THM MCL were 2.7-mg/L as C DOC (88%), 156 µg/L TOXFP (80%), and 3 cpu color (98%). The inorganic water quality was 145-mg/L TDS (65%), 68-mg/L Cl^- (32%), 85-mg/L as $CaCO_3$ TH (64%), and 7-mg/L as $CaCO_3$ Alk (58%). The pH was 7.8 and the water was stable.

A 25,000 gpd mobile UF pilot plant utilizing the N-50 membrane was built in a 30' trailer. The UF plant was housed in the 20x8' rear section of the trailer and equipped for antiscalant feed, acid feed, chlorination, stabilization, prefiltration, and storage as well as UF with variable recovery (50% to 90%) and feed pressure (80-120 psi). This plant was installed and operated at VOG for 365 hours from January 2 to March 3, 1985. Initially an operational test matrix was developed for water quality and flux from varying percentage of recovery and pressure. Product water DOC, color, and THMFP were independent of recovery and pressure over the test conditions and averaged less than 2-mg/L as C, 1 cpu, and 50 µg/L. Product water TDS and TH increased with increasing recovery, were independent of pressure, and varied from 25% to 75% of the raw water value. Operating conditions at VOG were set at 90 to 105 psi and 75% recovery in order to produce a water with a TH of 150-mg/L as $CaCO_3$, essentially no color, and THMFP of 50 µg/L or less. During the VOG operation, the product water quality and the percent rejection (% R) from the raw were 1.9-mg/L as C DOC (88%), 3 cpu color (97%), 27 µg/L THMFP, and 47 µg/L as Cl TOXFP. The raw water, TDS, TH, and Alk were reduced to 195-mg/L (60%), 142-mg/L as $CaCO_3$ (62%), and 135-mg/L as $CaCO_3$ (60%), respectively. The final pH was 7.5 and the water was stable. The product water flux declined 32% during the VOG operation from 20 to 13.4 gpd/ft². The water temperature was approximately 25°C and essentially did not vary during the operation.

On March 3, 1985, the UF pilot plant was moved from VOG to AID and operated until May 31, 1985, with an elapsed time of operation of 1020 hours. A second operational test matrix was developed that showed color, DOC, and THMFP removal was independent of recovery and pressure over the test conditions (50% to 90% product recovery and 80 to 120 psi). Product water color, DOC, and THMFP were typically less than 2 cpu, 2-mg/L as C, and 30 µg/L. TH and TDS removal was independent of pressure and dependent on product recovery and varied from 25% to 75% of the raw water value. Long-term operating conditions varied from 90-103 psi and 67-83% recovery. The product water quality and percent rejection (%R) from the raw were 1.5 cpu color (97%), 2.0-mg/L as C DOC (86%), 50 µg/L THMFP, and 48 µg/L as Cl TOXFP. The product water TDS, TH, and Alk values and percent rejection were 282-mg/L (43%), 187-mg/L as $CaCO_3$ (40%), and 180-mg/L as $CaCO_3$ (37%). The pH averaged 7.5 and the product was stable.

The UF pilot plant was designed with the membranes in four pressure vessels connected two each in series. The average pressure drop in the first pressure vessel was 32 psi (11 psi/membrane) and 41 psi (13 psi/membrane) in the second pressure vessel. The flux at AID rose slowly from 14 to 15 gpd/ft² during the first 150 hours of operation and after a chemical cleaning rose to slightly over 20 gpd/ft² and remained there for the duration of the study.

Cost of construction of a UF plant should be slightly less than an equivalent RO plant because of less expensive membranes and pressure requirements. Power costs should generally decrease from 75% (100 psi/400 psi) to 50% (100 psi/200 psi). The construction and O&M cost were estimated at $0.29/gal and $0.47/1000 gal for a 1 MGD UF plant.

6.5 Flotation

Flotation was one of the new technologies selected for investigation. The studies were conducted on a bench scale and analyzed for potential pilot testing. These experiments were conducted on site at Olga, utilizing a small flotation pilot that has been loaned to UCF and stored in a trailer on site. The unit has 1 ft² of surface area available for flotation and can be used either for pilot or bench scale testing. The flotation pilot plant was used in both capacities, but mainly as a source to generate white water (water supersaturated with dissolved air) that could be used in batch tests. A picture of the DAF unit in the trailer is shown in Figure A-15.

6.5.1 DAF

The results of three separate DAF studies using alum and PACl as coagulants are reported in Tables 114 and 115. These results show that DAF of alum or PACl sludges is very difficult and likely impractical. The main problem was the post precipitation of alum sludge after sample collection. The aluminum residuals as mg/L Al in these samples ranged from 1.3 to 7.8-mg/L as shown in Table 114. This residual is from alum that did post-precipitate after sample collection and was redissolved during acidification prior to aluminum analysis. These levels would cause problems with consumer acceptance and possibly consumer health. The maximum DOC removed during DAF tests 1 and 2 ranged from 65% for alum to 60% for PACl. These removals are no more than was obtained by conventional jar testing and indicate that DAF offers no advantage over coagulation for DOC (THM precursor) removal. The DOC measurements were made on filtered samples without disturbing the post precipitated aluminum sludge and therefore represent maximum DOC removal for that test.

However, additional DAF studies were conducted in a new reactor to avoid floc shearing during transfer of the coagulated sample to the DAF cylinder. The new reactor was a 8" diameter plexiglass cylinder with 4 stators running parallel to the 18" cylinder walls. Coagulation was carried out in the reactor as well as flotation. A whitewater distribution nozzle was constructed and installed in the bottom of the reactor to avoid floc shear. The results from these studies are shown in Tables 115. A picture of this reactor is shown in Figure A-16.

The data in Table 115 show the results of the in situ coagulation and DAF. These results are not appreciatively different than the other DAF studies. DOC removal is lower, 50% compared to 65%. There was post precipitation in the collected samples again and color removal was also less than would be expected by equivalent alum doses in conventional jar testing. Based on these data and the absence of existing literature concerning flotation of coagulated alum, it can be concluded that alum flotation by DAF offers no advantage for DOC removal relative to conventional coagulation and, as studied, destroys alum floc causing post precipitation.

6.5.2 Surfactants

Tests investigating frothers supplied by American Cyanamid for organic reduction from a high color, low turbidity, potable water source were completed. Two frothers studied were both basic alcohols and are marketed under the trade names, Aerofroth 76 and Aerofroth 65. The results of these bench tests are shown in Table 116. Neither surfactant was beneficial to any aspect of water treatment with color and DOC both increasing after froth flotation was completed. Although the data in Table 116 appears to suggest that both frothers reduced the THM's, it should be noted that the THM's measured were as a result of chlorination with a chlorine dose of 20-mg/L Cl_2 and no free residual was available after 48 hours in any sample. While the long chain alcohols that are surface active may have different THM precursor characteristics there is no demonstrated benefit to

Table 114. Olga DAF Tests 1 and 2

Sample	Coagu-lation	Dose mg/L	Polymer	Dose mg/L	pH	R	Color		Turbidity NTU	Al mg/L	DOC mg/L	Float mL
							Apparent cpu	True cpu				
Raw	--	--	--	--	7.8	--	90	90	2.0	0.2	19.2	--
Flotation	--	100	Plant	1	7.2	1	55	11	8.0	2.4	8.8	200
Flotation	Alum	100	Plant	1	7.2	2	63	11	4.2	3.6	9.0	200
Flotation	Alum	100	Plant	1	7.3	2	37	21	3.1	3.1	9.8	200
Flotation	Alum	200	Plant	1	5.8	1	63	7	7.6	5.3	8.6	250
Flotation	Alum	200	Plant	1	5.9	1	51	25	5.6	3.6	8.6	400
Flotation	Alum	200	Plant	1	5.8	2	20	13.5	2.6	1.6	6.8	240
Raw	--	--	--	--	7.8	--	110	85	2.2	.2	17.8	--
Flotation	Alum	100	Plant	1	6.5	1	110	35	8.7	7.4	9.5	280
Flotation	Alum	100	Plant	1	6.4	2	80	17	6.6	6.0	9.9	240
Flotation	Alum	200	Plant	1	5.8	1	75	46	9.1	7.6	7.2	380
Flotation	Alum	200	Plant	1	5.8	2	65	7	7.8	6.2	6.3	260
Raw	--	--	--	--	8.2	--	250	64	4.8	1.3	18.4	--
Flotation	PACl	100	Plant	1	6.8		45	12	4.0	4.3	9.2	400
Flotation	PACl	100	Plant	2	6.8		90	15	7.0	6.6	9.5	280
Flotation	PACl	200	Plant	1	6.4		69	9	6.5	6.3	7.9	320
Flotation	PACl	200	Plant	2	6.4		75	30	8.5	7.8	7.4	260

Comment: Flocculant particles were broken up during transfer from jar to graduated cylinder and when white water was injected.

R is the ratio of the coagulated water divided by the white or air saturated water.

Table 115. Olga DAF Test 3

Sample	Coagulation	Dose mg/L	Polymer	Dose mg/L	pH	R	Color Apparent cpu	Color True cpu	Turbidity NTU	Al mg/L	DOC mg/L	Float mL
Raw	--	--	--	--	8.0	--	80	63	2.2		18.3	--
Whitewater	--	--	--	--	10.3	--	51	22	3.1		11.0	--
Coagulation	Alum	50	A-243	0.5	7.3	--	110	29	7.1		15.1	--
Coagulation	Alum	100	A-243	0.5	6.9	--	32	15	2.6		13.1	--
Coagulation	Alum	150	A-243	0.5	6.75	--	33	12	3.0		10.3	--
Coagulation	Alum	200	A-243	0.5	6.45	--	38	8	4.6		9.4	--
Flotation	Alum	100	A-243	0.5	7.9	1	80	21	9.9		12.4	0
Flotation	Alum	100	C-2466	0.5	7.8	1	97	24	9.3		13.0	0
Flotation	Alum	100	A-2706	0.5	7.8	1	85	23	9.7		11.8	0
Flotation	Alum	200	N-2696	0.5	7.1	1	100	13	10.5		10.4	550
Flotation	Alum	200	Amp-2996	0.5	7.1	1	105	12	12.5		10.0	200
Flotation	Alum	200	A-243	0.5	7.0	1	105	13	12.9		11.1	100
Flotation	Alum	200	C-2466	0.5	7.0	1	105	14	12.7		10.2	100
Flotation	Alum	200	A-2706	0.5	7.0	1	100	15	12.1		9.5	150
Flotation	Alum	200	N-2696	0.25	7.0	1	105	12	12.6		9.5	100
Flotation	Alum	200	N-2696	1.0	7.1	1	60	11	8.6		10.5	1200
Flotation	Alum	200	N-2696	2	7.1	1	16	10	2.2		10.0	1000
Flotation	Alum	200	N-2696	5	7.0	1	17	9	2.3		9.4	675

Table 116. Test Data for Frother Addition to Olga Raw Water

Sample	Frother Dose	pH	DOC	Color Apparent	THM's[1]
	mg/L		mg/L	CPU	mg/L
Raw		7.7	18.2	75	1198
A-76	20	8.3	28.1	64	1314
A-76	25	8.3	30.9	56	934
A-76	50	8.4	43.9	68	999
A-65	20	8.3	29.1	71	970
A-65	25	8.3	33.1	73	992
A-65	50	8.4	47.1	66	NA

Frothers:

(A-76) = Aerofroth 76 American Cyanamid Co.

(A-65) = Aerofroth 65 American Cyanamid Co.

[1] THM's measured by dosing 20 mg/L Cl_2 at 22°C, and pH 7.6 for 48 hrs.

water treatment for removal of THM precursors. Therefore these studies were completed with the conclusion that neither DAF nor froth floatation was beneficial for THM precursor removal for the conditions utilized in this study. However, the Caloosahatchee River is a low turbidity, high color source. If a high turbidity water was investigated, some advantage of flotation might be demonstrated.

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APPENDIX

Table A-1. AID - Process pH Optimization Study

Acme - Jar Test

Sample	Reaction pH	DOC (mg/L as C)	TOXFP* (mg/L Cl)	THM's* (µg/L)	Color (cpu)	TH (mg/L as $CaCO_3$)	CaH (mg/L as $CaCO_3$)	Cond. (mmhos)	pH
Acme -	Raw	12.1	1575	475	29	314	308	.654	7.60
Acme -	9.0	11.3	1359	415	26	232	194	.471	9.00
Acme -	9.5	10.6	1075	317	18	90	70	.314	9.40
Acme -	10.0	9.2	957	295	22	67	44	.283	10.00
Acme -	10.3	9.4	1308	268	22	55	48	.288	10.30
Acme -	10.7	9.2	1032	350	14	51	50	.292	10.60
Acme -	11.0	8.8	907	252	11	89	84	.369	11.00
Acme -	11.5	7.8	833	341	11	169	137	.958	11.50

*Note: THM and TOX are 48 hr. formation potential values, incubation at 20°C and pH = 7.5

Reaction pH raised with $Ca(OH)_2$ and adjusted with H_2SO_4 if necessary.

Table A-2. Village of Golf - Process pH Optimization Study

VOG - Jar Test

Sample	Reaction pH	DOC (mg/L as C)	TOXFP* (μg/L Cl)	THM's* (μg/L)	Color (cpu)	TH (mg/L as $CaCO_3$)	CaH (mg/L as $CaCO_3$)	Cond. (mmhos)	pH
VOG -	Raw	12.32	1849	697	60	235	218	0.503	8.71
VOG -	9.0	12.03	1804	513	48	170	156	0.427	9.00
VOG -	9.5	11.57	1816	614	48	118	100	0.418	9.50
VOG -	10.0	11.38	1742	501	37	34	32	0.312	10.00
VOG -	10.3	11.74	1815	432	37	76	74	0.327	10.30
VOG -	10.7	10.90	1576	388	30	74	72	0.348	10.60
VOG -	11.0	9.86	1465	408	22	88	77	0.418	11.00
VOG -	11.5	7.49	1068	428	11			1.479	11.50

*Note: THM and TOX are 48 hr. formation potential values, incubation at 30° C and pH = 7.5

Reaction pH raised with $Ca(OH)_2$ and adjusted with H_2SO_4 if necessary.

Table A-3. Village of Golf - Coagulant Optimization Study Results

Sample	DOC (mg/L as C)	THM's (μg/L)	Color (cpu)	Cond. (mmhos)	Al (in mg/L)
AC-190 (0.8)[a]	8.59	397	16	0.710	0.52
AC-190 (1.6)	8.68	340	20	0.705	1.21
AC-190 (2.4)	8.41	233	23	0.708	2.11
AC-190 (4.0)	8.23	269	16	0.561	3.16
AC-100S (0.8)[b]	8.18	255	16	0.930	0.61
AC-100S (1.6)	8.20	317	20	0.923	1.54
AC-100S (2.4)	8.42	353	20	0.854	2.23
AC-100S (4.0)	7.96	321	17	0.854	4.34
Alum (0.8)[c]	7.87	326	16	0.781	0.29
Alum (1.6)	8.45	321	9	0.873	0.65
Alum (2.4)	7.93	332	27	0.934	1.07
Alum (4.0)	8.09	323	16	0.910	2.15
Control[d]	9.49	328	15	--	--

[a] PACl 190 Coagulant - Number in parenthesis refers to initial Al dose in mg/L

[b] PACl 100S - Number in parenthesis refers to initial Al dose in mg/L

[c] Alum - Commercial Alum - Number in parenthesis refers to initial Al dose in mg/L

[d] THM values are 48 hr formation potentials at 20°C with an initial Cl_2 dose of 26 mg/L and a pH of 7.60

Table A-4. Village of Golf - 30 Minute Cl_2 Induced Color Removal

Plant Chemical Dosages		UCF Chemical Dosages	
Cl_2 Dose in mg/L Cl_2	Color in cpu	Cl_2 Dose in mg/L Cl_2	Color in cpu
0	42	0	42
1.0	38	1.0	29
2.5	29	2.5	17
5.0	27	5.0	14
7.5	24	7.5	12
10.0	20	10.0	7
15.0	15	15.0	8
20.0	13	20.0	5
30.0	15	30.0	2

Table A-5. Village of Golf Chlorine Dissipation vs. Time Study (all values in mg/L as Cl_2)

Sample	Time (in hours)					
	1	4	24	48	72	96
VT-20*	16.5	13.4	10.7	8.0	7.5	5.4
VT-10*	6.2	4.0	0.9	0	0	0
VT-5*	0.3	0	0	0	0	0
VP-26*	NA	18.8	12.1	10.7	7.2	5:4

VT = Village of Golf - Optimized Chemical Dosages
VP = Village of Golf - Plant Chemical Dosages

* Number refers to Initial Chlorine Dose in mg/L as Cl_2

Table A-6. Village of Golf - THM Formation using Typical Plant Chemical Doses* at 30° C and pH 8.8

Time in hours	THM's in μg/L
Instant	104
1	192
5	401
24	857
48	953
72	1060
96	1112

*Lime = 175 mg/L as $Ca(OH)_2$
Soda Ash = 75 mg/L as Na_2CO_3
Polymer = 0.7 mg/L
Chlorine = 26 mg/L as Cl_2

Table A-7. Village of Golf - THM Formation using UCF Experimental Chemical Doses* at 30° C and pH 8.8

Time in hours	THM's in μg/L
Instant	34
1	107
10	295
24	402
48	534
72	642
96	

*Alum = 0.80 mg/L as Al^{3+}
Lime = 350 mg/L as $Ca(OH)_2$
Soda Ash = 75 mg/L as Na_2CO_3
Polymer = 0.70 mg/L Polyhall
Chlorine = 15 mg/L as Cl_2

TABLE A-8. Color Throughout the Plant During the Dry Season

Date	Raw	Softened	Color (cpu) Coagulated	Transfer Basin	Tap
Feb. 7, 1985	64	18	11.0	12.0	6.0
" 8, "					
" 9, "	64	25	18.5	17.5	6.5
" 10, "	66	19	10.5	9.0	3.5
" 11, "	68	44	14.5	7.5	7.0
" 12, "					
" 13, "	63	39	11.0	3.0	3.5
" 14, "					
" 15, "	63	33	12.0	3.0	2.5
" 16, "					
" 17, "					
" 18, "					
" 19, "	60	25	5.0	6.0	2.0
" 20, "					
" 21, "	62	28	5.5	2.0	2.0
" 22, "	59	30	7.0	2.0	1.5
" 23, "	63	29	8.0	1.0	3.0
" 24, "	58	33	9.0	1.0	2.0
" 25, "	60	29	12.0	1.5	2.0
" 26, "	57	40	7.5	2.0	2.5
" 27, "	58	40	10.5	3.0	2.0
" 28, "	56	36	8.0	1.6	2.0
March 1, 1985	56	34	8.0	2.0	2.0
" 2, "	59	32	14.5	4.0	2.0
" 3, "	58	33	16.0	3.5	3.0
" 4, "	56	30	11.0	3.5	3.5
" 5, "	55	27	13.0	6.0	4.0
" 6, "	55	35	12.0	4.0	4.0
" 7, "					
" 8, "					
" 9, "	54	28	1.0	2.0	1.0
" 10, "	52	32	17.0	3.0	3.0
" 11, "	56	33	11.0	3.0	5.0
" 12, "	56	38	15.0	4.0	4.0
" 13, "	54	35	15.0	4.0	4.0
" 14, "	53	38	17.0	6.0	6.0
" 15, "	54	28	14.0	3.0	4.0
" 16, "	53	25	8.0	2.0	2.0
" 17, "	50	21	6.0	1.0	1.0
" 18, "	53	18	11.0	5.0	6.0
" 19, "	56	17	8.0	7.0	6.0

TABLE A-9. Olga Plant Tests Aluminum Residuals, Lee County Data (mg/L)

Date	1	2	3	4	5	6	7	8
2-8-85	1.14	0.91	1.11	0.77	1.05	0.85	1.05	1.17
2-9-85	0.88	0.66	1.97	0.97	1.26	1.05	0.91	1.77
2-10-85	1.02	0.87	1.48	0.95	1.15	1.18	0.90	1.69
2-11-85	0.83	0.45	0.69	0.66	1.08	1.19	1.05	0.60
2-12-85	0.63	0.39	0.57	1.72	1.05	0.90	1.01	
2-13-85	0.39	0.30	0.42	0.51	0.81	0.78	1.10	0.54
2-14-85	0.26	0.31	0.16	0.35	0.43	0.26	0.36	0.21
2-15-85	0.47	0.14	0.02	0.05	0.40	0.26	0.36	0.05
2-16-85	0.14	0.05	0.05	0.07	0.05	0.12	0.22	0.03
2-17-85	0.09	0.57	2.08	0.33	0.30	0.02	0.09	1.97
2-18-85								
2-19-85	1.73	1.78	1.56	1.78	1.63	1.70	1.49	1.68
2-20-85	1.52	0.78	0.10	0.56	1.19	1.65	1.78	0.10
2-21-85	0.60	0.40	0.15	0.50	0.30	0.30	0.35	
2-22-85	0.25	0.20	0.10	0.10	0.25	0.25	0.20	0.10
2-23-85	0.30	0.25	0.10	0.15	0.10	0.10	0.25	0.10
2-24-85	0.15	0.15	0.10	0.10	0.15	0.10	0.20	0.10
2-25-85	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10
2-26-85	0.20	0.10	0.20	0.10	0.10	0.10	0.20	0.10
2-27-85	0.30	0.10	0.10	0.10	0.10	0.10	0.20	0.10
2-28-85	0.20	0.10	0.15	0.15	0.10	0.10	0.15	0.10
3-1-85	0.10	0.10	0.10	0.10	0.20	0.10	0.10	0.10
3-2-85	0.10	0.15	0.10	0.20	0.10	0.10	0.10	0.10
3-3-85	0.10	0.15	0.10	0.15	0.15	0.20	0.10	0.10
3-4-85	0.10	0.10	0.10	0.10	0.15	0.20	0.10	0.10
3-5-85	0.15	0.15	0.30	0.15	0.15	0.10	0.20	0.10
3-6-85	0.15	0.25	0.10	0.10	0.20	0.10	0.10	0.10
3-7-85	0.25	0.15	0.20	0.20	0.10	0.10	0.10	0.10
3-8-85	0.15	0.15	0.30	0.10	0.10	0.15	0.10	0.15
3-9-85	0.10	0.10	0.10	0.15	0.10	0.10	0.10	0.10
3-10-85	0.10	0.10	0.10	0.15	0.10	0.15	0.10	0.10
3-11-85	0.10	0.10	0.15	0.10	0.10	0.10	0.10	0.10
3-12-85	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10
3-13-85	0.10	0.15	0.20	0.10	0.10	0.10	0.15	0.25
3-14-85	0.25	0.15	0.20	0.10	0.20	0.10	0.10	0.20
3-15-85	0.10	0.15	0.15	0.15	0.15	0.10	0.10	0.10
3-16-85	0.10	0.15	0.10	0.10	0.10	0.10	0.10	0.10
3-17-85	0.10	0.10	0.10	0.10	0.10	0.10	0.15	0.10
3-18-85	0.10	0.10	0.35	0.10	0.15	0.10	0.15	0.10

TABLE A-10. Olga Alum/PACl Plant Data Wet Season

Date	Coagulation			DOC		Color		Inst THM	TOXFP	Al
	Type	mg/L	pH	mg/L	%R	cpu	%R	ug/L	ug/L	mg/L
July 20, 1985										1.29
21										1.03
22										1.18
23										1.48
24	Alum	190	7.8	15.1	48	80	49	605	1173	1.74
25	"	230	6.8	8.8	57	16	89	505	--	0.59
26	"	320	5.6	5.7	74	9	95	395	740	0.28
27	"	165	5.6	5.8	73	7	96	410	--	0.18
28	"	190	5.8	6.3	72	8	95	250	634	0.17
29	"	230	5.9	6.5	74	8	95	415	712	0.27
30	"	230	5.9	7.0	70	10	94	364	--	0.13
31	"	230	5.7	7.1	71	7	96	340	782	0.17
Aug. 1 1985	"	230	5.7	6.5	72	8	95	141	718	0.30
2	"	230	5.6	6.6	69	8	95	215	--	0.29
3	"	160	6.1	7.6	62	12	92	225	--	0.10
4	"	160	6.2	1.6	46	22	85	203	--	0.20
5	"	160	6.4	8.4	63	26	84	181	--	0.23
6	"	160	6.3	--	--	--	--	183	--	0.16
7	"	160	6.2	8.2	62	13	91	182	--	0.13
8	"	160	5.9	--	--	2	98	160	--	0.10
9	"	160	5.9	7.3	64	1	99	147	--	0.17
10	"	160	5.6	9.4	59	2	98	199	--	0.18
11	"	80	5.7	7.6	67	2	98	155	--	0.18
12	"	80	5.8	9.2	57	11	94	187	--	0.18
13	"	80	5.7	9.4	57	8	94	180	561	0.18
14	"	80	5.7	9.6	55	11	93	180	776	0.15
15	"	230	5.5	6.5	68	6	95	229	891	0.19
16	PACl	80	5.8	8.3	62	9	92	195	578	0.21
17	"	160	5.8	7.9	63	8	93	234	838	0.13
18	"	160	5.8	8.6	60	10	91	238	205	0.19
19	"	160	5.8	10.0	50	14	87	148	626	0.24
20	"	160	6.8	9.2	50	15	86	329	574	0.11
21	"	120	6.2	8.4	51	12	91	236	638	0.27
22	"	120	6.1	8.5	57	14	89	209	842	0.23
23	"	120	6.0	9.8	50	13	88	351	917	
24	"	240	5.8	7.5	62	6	96	187	766	0.16
25	"	240	5.8	7.3	65	6	96	225	675	0.13
26	"	160	6.2							
27	"	160	6.2							

%R = percent removal from raw

Table A-11. Trihalomethanes (μg/L) in Distribution System During the Wet Season

	Site							
Date	1	2	3	4	5	6	7	8
07/20/84	--	965	853	917	1,037	1,021	1,055	756
07/24/84	837	813	688	763	852	748	---	---
07/25/84	896	776	581	776	929	938	---	559
07/26/84	253	643	854	259	692	944	---	559
07/27/84	315	593	500	524	704	697	---	445
07/28/84	713	391	465	496	635	649	---	471
07/29/84	510	534	448	587	512	508	---	415
07/30/84	437	398	384	416	523	523	---	364
07/31/84	521	466	430	438	477	460	---	347
08/02/84	389	306	304	334	450	334	---	237
08/04/84	298	288	208	275	357	322	363	188
08/06/84	249	275	202	211	300	286	323	183
08/08/84	254	226	189	228	229	321	332	193
08/10/84	292	243	184	258	305	284	298	196
08/11/84	276	266	285	308	262	305	324	248
08/12/84	318	286	424	284	262	296	586	216
08/13/84	337	296	258	315	313	329	316	215
08/14/84	315	317	209	286	290	295	309	216
08/15/84	295	280	247	324	319	318	335	228
08/16/84	273	327	246	267	341	348	336	273
08/17/84	331	344	345	326	392	380	369	344
08/18/84	300	369	278	375	360	368	338	262
08/19/84	308	325	305	289	388	342	367	197
08/20/84	284	283	263	301	292	407	440	294
08/21/84	319	359	277	395	421	399	390	290
08/22/84	322	412	259	373	394	390	371	236
08/23/84	427	385	284	328	372	412	404	279
08/24/84	330	326	251	307	306	299	331	257
08/25/84	323	291	194	327	309	319	350	162
08/26/84	347	255	216	229	335	353	377	194
08/27/84	259	264	158	239	376	340	350	287
08/28/84	332	334	263	310	307	257	405	247

Location
1 - Orange River Lift Station
2 - 4 M Lift Station
3 - Old Olga Lift Station
4 - West Bayshore Lift Station
5 - North Reservoir Storage Tank
6 - 438 Eland (Furtherst Pt.)
7 - Six Lake Mobile Home Park
8 - Plant Tap

TABLE A-12. Plant Sequential Treatment Data Following Lime Softening

Date	Lime mg/L CaO	pH	TH mg/L $CaCO_3$	DOC mg/L	% Reduction	Color cpu	% Reduction	THMFP ug/L	TOXFP ug/L
3-29	210	10.3	120	10.52	51	19	57	472	877
3-30	210	10.2	118	9.88	54	19	57	379	899
3-31	215	10.3	118	10.32	39	15	67	312	852
4-5	225	10.4	140	10.35	38	20	56	273	962
4-6	225	10.4	130	9.81	40	13	70	215	823
4-6	225	10.4	130	9.90	36	14	67	370	764
4-13	210	10.3	116	9.38	39	20	58	274	847
4-13	225	10.4	110	10.96	34	21	55	295	848
4-14	210	10.3	90	9.93	39	20	57	259	751
4-14	210	10.2	110	10.54	41	22	55	323	847
4-15	210	10.3	90	10.67	33	22	55	305	790
4-15	210	10.3	105	11.54	30	24	50	339	741
4-16	210	10.3	110	8.97	41	20	59	402	522
4-20	210	10.4	120	8.98	39	17	58	400	714
4-21	210	10.3	100	9.13	40	19	58	472	697
5-3	250	11.0	113	7.78	52	12	74	330	720
5-4	270	11.3	217	7.66	52	11	76	447	743
5-5	250	10.9	176	9.12	43	18	60	573	832
5-5	260	11.2	187	9.45	44	12	70	429	876

TABLE A-13. Plant Sequential Treatment Data Following Alum Coagulation

Date	Alum mg/L $Al_2(SO_4)_3 \cdot 14H_2O$	pH	TH mg/L $CaCO_3$	DOC mg/L	$\%R_1$	$\%R_2$	Color CPU	$\%R_1$	$\%R_2$	THMFP mg/L	$\%R_1$	TOXFP mg/L	$\%R_1$
3/29	68	7.1	140	9.14	13	57	8	58	87	330	30	849	3
3/30	68	7.1	110	6.15	38	71	4	79	91	199	47	853	5
3/31	68	7.9	120	4.10	12	46	13	13	71	286	8	770	10
4/5	76	6.6	106	7.04	32	58	7	65	84	229	16	637	34
4/6	76	6.4	140	7.10	28	56	6	54	86	145	33	735	11
4/6	76	6.6	140	8.35	16	46	12	14	72	228	38	696	9
4/13	83	6.2	118	5.97	36	61	10	50	79	209	24	489	42
4/13	83	6.0	100	5.55	49	66	7	67	85	277	6	525	38
4/14	83	6.1	108	5.74	42	65	7	65	85	204	21	400	47
4/14	94	5.3	98	5.97	44	67	7	68	86	222	33	480	42
4/15	94	5.3	98	5.80	46	63	7	68	86	263	14	559	29
4/15	94	5.5	94	5.95	48	64	7	71	85	299	12	501	32
4/16	109	5.0	102	4.73	47	68	2	82	93	375	6	550	23
4/20	109	5.0	102	4.73	47	68	2	82	93	365	6	550	29
4/21	109	5.0	110	4.20	54	72	3	84	93	336	29	606	13
5/3	109	6.5	159	4.31	45	73	3	75	93	231	30	471	35
5/4	109	6.3	227	4.93	36	69	5	55	85	293	34	613	17
5/5	109	5.5	172	4.64	49	71	4	78	91	316	45	593	29
5/5	109	5.0	149	4.75	60	72	4	82	90	414	3	643	27

$\%R_1$ = percent removal relative to the softened water

$\%R_2$ = percent removal relative to the raw water

Table A-14. Olga N-50 Ultrafiltration Study - Removals Based on Feed Concentrations.

Date	Hours of Operation	Sample Type/% % Removal	Feed Pressure (psi)	Percent Recovery (%)	DOC (mg/L as C)	Color (cpu)	TH (mg/L as $CaCO_3$)	TDS (mg/L)	Chlorides (mg/L)
4/11	5.5	Feed	75	60	38.24	129	425	809	160
		Product			1.75	2.5	69	157	80
		% Removal			95.4	98.1	83.8	80.6	50.0
4/15	90	Feed	75	75	68.17	N.A	631	1016	191
		Product			2.44	3.0	86	186	95
		% Removal			96.4		86.4	81.7	50.3
*4/18 (1)	151	Feed	75	60	46.35	N.A.	398	720	126
		Product			1.65	2.0	69	144	60
		% Removal			96.4		82.7	80.0	52.4
4/18 (2)	154	Feed	75	60	50.00	N.A.	376	720	117
		Product			1.44	2.5	75	137	66
		% Removal			97.1		80.1	81.0	44.4
4/18 (3)	157	Feed	75	60	46.57	N.A.	408	715	129
		Product			1.37	1.5	79	141	66
		% Removal			97.1		80.1	80.3	48.8
4/23	218	Feed	95	68	47.27	N.A.	463	1042	166
		Product			2.54	5.0	69	171	77
		% Removal			94.7		85.1	83.6	53.6
4/25 (1)	263	Feed	95	72	50.29	N.A.	481	N.A.	175
		Product			2.71	4.0	93	258	89
		% Removal			94.6		80.1		49.1
4/25 (2)	271	Feed	60	61	43.73	N.A.	418	694	172
		Product		2.0	2.01	3.0	76	200	67
		% Removal		89.8	95.4		81.8	71.2	61.0

* Membrane cleaning 4/18 (3 samples)
N.A. - Not Available

(continued)

Table A-14 (continued). Olga N-50 Ultrafiltration Study - Removals Based on Feed Concentrations

Date	Hours of Operation	Sample Type/% % Removal	Feed Pressure (psi)	Percent Recovery (%)	DOC (mg/L as C)	Color (cpu)	TH (mg/L as $CaCO_3$)	TDS (mg/L)	Chlorides (mg/L)
5/2	342	Feed	65	90	119.50	289	683	1228	190
		Product			12.76	26	163	224	86
		% Removal			89.3	91.0	76.1	72.7	54.7
5/6 (1)	426	Feed	70	75	71.87	130	512	952	123
		Product			6.13	10.0	119	228	72
		% Removal			91.5	92.3	76.8	76.1	41.5
5/6 (2)	435	Feed	63	60	53.96	130	399	735	114
		Product			4.85	6.0	93	208	66
		% Removal			91.0	95.4	76.7	71.7	42.1
5/7	442	Feed	105	60	53.85	123	374	714	124
		Product			3.76	4.0	81	162	61
		% Removal			93.0	96.8	78.3	77.3	50.8
5/9	498	Feed	105	60	46.31	115	394	650	125
		Product			2.56	3.5	73	138	58
		% Removal			94.5	97.0	81.5	78.8	53.6
5/13	590	Feed	105	60	51.69	134	411	767	134
		Product			2.66	4.0	78	154	68
		% Removal			94.8	97.0	81.0	80.0	49.2
5/16	680	Feed	90	60	50.10	120	393	772	129
		Product			2.04	4.0	75	170	70
		% Removal			95.9	96.7	80.9	78.0	45.7
5/17	703	Feed	90	58	N.A.	123	360	738	146
		Product			4.06	7.0	94	201	79
		% Removal				94.3	73.9	72.8	45.9

Table A-15. Olga N-50 Ultrafiltration Study (Inorganics)

Date (HR)	Sample Type/ % Removal	Feed Pressure (psi)	Percent Recovery (%)	pH	Alk (mg/L as $CaCO_3$)	TH (mg/L as $CaCO_3$)	CaH (mg/L as $CaCO_3$)	TDS (mg/L)	Chlorides (mg/L)	Sodium (mg/L)
4/11 (5.5)	Raw	75	60	8.1	176	239	186	424	120	74
	Product			7.7	50	69	56	157	80	46
	% Removal				71.5	71.1	69.9	63.0	33.0	37.8
4/15 (90)	Raw	75	75	7.9	174	242	180	474	122	N.A
	Product			7.7	94	86	80	186	95	59
	% Removal				46.0	64.6	55.5	60.8	22.1	N.A
4/18 * (BC) (151)	Raw	75	60	8.0	176	215	175	425	93	60
	Product			7.7	63	69	54	144	60	40
	% Removal				64.2	67.9	69.1	66.1	35.5	33.3
4/18 (AC) 1 hr (154)	Raw	75	60	8.0	176	215	175	425	93	60
	Product			7.8	68	75	55	137	65	39
	% Removal				61.4	65.1	68.6	67.8	30.1	35.0
4/18 (AC) 4 hr. (157)	Raw	75	60	8.0	176	215	175	400	93	60
	Product			7.9	66	79	72	171	65	40
	% Removal				62.5	63.3	58.9	57.2	30.1	3.3
4/23 (218)	Raw	95	68	8.1	155	206	175	491	119	65
	Product			7.9	71	69	67	258	77	40
	% Removal				54.2	65.5	61.7	47.4	35.5	38.4
4/25 (263)	Raw	90	70	8.2	149	218	180	491	116	65
	Product			7.8	84	93	74	258	89	58
	% Removal				43.6	57.3	58.8	47.4	23.2	10.7
4/25 (271)	Raw	60	61	8.1	160	222	180	484	119	N.A
	Product			7.8	67	76	61	200	67	N.A
	% Removal				58.1	65.8	66.1	58.7	43.7	N.A

*Membrane cleaned with solution of EDTA and NaOH at pH 11.0 per Filmtec procedure
BC - Before cleaning   AC - After cleaning

(continued)

Table A-15 (continued). Olga N-50 Ultrafiltration Study - Bench Scale Test Results (Inorganics)

Date (HR)	Sample Type/ % Removal	Feed Pressure (psi)	Percent Recovery (%)	pH	Alk (mg/L as $CaCO_3$)	TH (mg/L as $CaCO_3$)	CaH (mg/L as $CaCO_3$)	TDS (mg/L)	Chlorides (mg/L)	Sodium (mg/L)
5/2 (342)	Raw			8.0	158	245	230	404	102	56
	Product	70	90	7.7	106	163	157	335	86	47
	% Removal				32.9	33.5	31.7	17.1	15.7	16.1
5/6 (1) (426)	Raw			8.3	160	221	170	387	91	48
	Product	63	60	7.7	80	93	85	208	66	40
	% Removal				50.0	57.9	50.0	46.3	27.5	16.7
5/6 (2) (435)	Raw			8.3	160	221	170	387	91	48
	Product	70	73	8.0	99	119	110	228	72	44
	% Removal				38.1	46.2	35.3	41.0	20.9	8.3
5/7 (442)	Raw			8.2	160	221	170	387	91	49
	Product	105	60	7.7	64	81	62	162	61	36
	% Removal				60.0	63.3	63.5	58.1	33.0	26.5
5/9 (498)	Raw			8.3	166	221	170	387	91	49
	Product	105	60	7.6	68	72	56	138	57	36
	% Removal				59.0	66.8	68.2	63.3	37.4	22.0
5/13 (590)	Raw			8.1	158	215	166	369	98	52
	Product	105	63	7.8	69	78	68	154	67	39
	% Removal				56.3	63.7	59.0	58.3	31.6	25.0
5/16 (680)	Raw			8.2	166	212	172	420	96	52
	Product	90	60	7.6	76	75	68	170	69	42
	% Removal				54.2	64.6	60.5	59.5	28.1	19.2
5/17 (703)	Raw			8.1	165	215	177	399	105	52
	Product	90	58	7.8	55	94	75	201	79	47
	% Removal				66.7	56.3	57.6	49.6	24.8	9.6

Table A-16. VOG Test Matrix - Percent Parameter Removal Based on Raw Water.

Feed Pressure (psi)	Product Recovery Rate (%)	Total Dissolved Solids (%)	Total Hardness (%)	True Color (%)
80	50	58	54	99
80	66	43	43	99
80	75	41	41	99
80	90	25	25	97
90	50	60	60	99
90	66	50	50	99
90	75	48	48	99
90	90	44	44	99
100	50	61	61	99
100	66	54	54	99
100	75	51	51	98
100	90	46	46	99
110	50	57	57	99
110	66	38	38	98
110	75	42	42	99
110	90	37	37	99

Table A-17. VOG Pilot Plant Operating Data Summary

Date	Time of Operation (hr:min)	Elapsed Time of Operation (hr:min)	Product Flux gals./day-ft²	Feed Pressure (psi)
12-27	3:00	3:00	19.6	92
1-2	3:35	6:35	20.0	90
1-3	6:35	13:10	20.0	91
1-4	6:30	19:40	19.5	92
1-5	6:50	26:30	19.4	91
1-6	5:10	31:40	19.4	92
1-7	7:25	39:05	19.4	91
1-8	7:05	46:10	19.2	91
1-9	4:30	50:46	19.2	91
1-10	4:35	58:15	19.2	91
1-11	7:00	65:15	19.2	91
1-12	6:40	72:05	19.0	93
1-13	5:35	78:40	19.0	91
1-14	7:30	86:10	18.9	92
1-15	7:15	93:25	18.6	91
1-16	5:50	99:15	18.4	91
1-17	6:35	105:50	NA	NA
1-18	7:00	112:50	NA	NA
1-19	3:25	116.15	NA	NA
1-20	5:50	122:05	NA	NA
1-21	7:20	129:25	NA	NA
1-22	2:40	132:05	NA	NA
1-23	6:10	138:15	NA	NA
1-24	7:35	145:56	NA	94
1-25	5:45	151:35	17.3	94
1-26	3:35	155:10	17.4	94
1-27	6:55	162:05	17.6	95
1-28	7:30	169:35	17.3	95
1-29	7:30	177:05	16.3	96
1-30	5:55	183:00	16.5	96
1-31	6:45	189:45	NA	95
2-1	0	189:45	16.0	NA
2-2	7:15	197:00	NA	101
2-3	0	197:00	15.8	NA
2-4	4:20	201:20	15.4	102
2-5	8:25	209:45	15.5	103
2-6	5:40	215:25	15.5	103
2-7	6:25	221:50	14.6	103
2-8	7:35	229:25	14.2	106
2-9	7:15	236:40	14.4	105
2-10	5:40	242:20	14.2	105
2-11	7:25	249:45	14.4	105
2-12	4:00	253:45	14.2	105
2-13	5:05	258:50	14.2	105

(continued)

Table A-17 (continued). VOG Pilot Plant Operating Data Summary

Date	Time of Operation (hr:min)	Elapsed Time of Operation (hr:min)	Product Flux gals./day-ft²	Feed Pressure (psi)
2-14	3:45	262:35	14.2	105
2-15	7:15	269:50	14.1	105
2-16	7:10	277:00	14.1	104
2-17	5:35	282:35	NA	NA
2-18	4:35	287:10	NA	NA
2-19	7:25	294:35	NA	NA
2-20	5:45	300:20	NA	NA
2-21	6:30	306:50	NA	NA
2-22	7:00	313:50	13.9	105
2-23	7:10	321:00	13.6	105
2-24	5:30	326:30	14.1	105
2-25	4:40	331:10	13.9	105
2-26	7:20	338:30	13.6	105
2-27	3:50	342:20	13.9	105
2-28	7:10	349:30	13.6	105
3-1	7:15	356:45	13.6	105
3-2	7:15	364:00	13.4	105
3-3	1:00	365:00	14.1	105

Table A-18. VOG Raw Water Quality Data Summary

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
1-5	29	91	75	13.8	60	225	247	355
1-8	48	91	75	14.3	58	234	277	357
1-11	67	91	75	15.3	56	227	284	413
1-14	87	92	74	15.6	59	219	228	146
1-17	107	NA	NA	15.8	56	221	224	361
1-20	123	NA	NA	17.1	62	229	229	313
1-23	139	NA	NA	18.1	63	225	223	343
1-26	156	94	71	15.1	56	229	244	383
1-29	178	96	69	13.8	55	220	236	366
2-2	197	101	74	17.2	59	227	229	353
2-7	222	103	71	16.9	58	222	264	361
2-10	242	105	69	17.2	58	226	260	379
2-13	259	105	68	16.6	58	236	245	388
2-16	277	104	68	15.2	58	225	251	196
2-10	295	NA	NA	NA	NA	NA	NA	NA
2-22	314	105	68	16.8	61	220	236	321
2-25	331	105	68	17.4	59	202	252	358
2-28	350	105	67	16.9	59	202	240	367
3-3	365	105	69	16.5	58	204	252	343

Table A-19. VOG Feed Water Quality Data Summary [Values from 1-5 to 1-29, inclusive, have been calculated from the raw and reject water quality-all other values are direct measurement].

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
1-5	29	91	75	31.4	120	315	347	517
1-8	48	91	75	31.4	120	322	370	659
1-11	67	91	75	31.9	120	318	375	560
1-14	87	92	74	32.1	120	307	335	548
1-17	107	NA	NA	NA	NA	NA	NA	NA
1-20	123	NA	NA	NA	NA	NA	NA	NA
1-23	139	NA	NA	NA	NA	NA	NA	NA
1-26	156	94	71	30.3	110	311	342	523
1-29	178	96	69	31.5	110	310	347	530
2-2	197	101	74	32.2	95	289	323	502
2-7	222	103	71	31.6	95	289	339	489
2-10	242	105	69	31.7	95	307	351	490
2-13	259	105	68	NA	NA	NA	NA	NA
2-16	277	104	68	25.2	95	309	316	512
2-10	295	NA	NA	NA	NA	NA	NA	NA
2-22	314	105	68	32.8	95	281	328	444
2-25	331	105	68	29.8	95	260	332	450
2-28	350	105	67	33.2	90	281	324	465
3-3	365	105	69	31.6	90	241	328	449

Table A-20. VOG Reject Water Quality Data Summary

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
1-5	29	91	75	64.0	230	482	534	817
1-8	48	91	75	63.2	240	484	544	1220
1-11	67	91	75	62.8	245	488	544	833
1-14	87	92	74	62.7	240	469	534	792
1-17	107	NA	NA	62.8	245	473	536	829
1-20	123	NA	NA	66.4	230	471	538	790
1-23	139	NA	NA	66.9	220	462	553	780
1-26	156	94	71	58.4	210	462	524	782
1-29	178	96	69	63.1	205	469	543	822
2-2	197	101	74	59.7	200	468	519	778
2-7	222	103	71	59.3	195	430	521	755
2-10	242	105	69	60.1	190	430	517	748
2-13	259	105	68	54.7	190	426	483	751
2-16	277	104	68	48.8	190	421	499	773
2-10	295	NA	NA	NA	NA	NA	NA	NA
2-22	314	105	68	58.2	205	418	480	693
2-25	331	105	68	49.9	190	372	469	703
2-28	350	105	67	54.6	195	378	473	696
3-3	365	105	67	49.1	195	378	488	685

Table A-21. VOG Product Water Quality Data Summary

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
1-5	29	91	75	1.6	1.0	135	144	207
1-8	48	91	75	1.7	2.0	142	138	215
1-11	67	91	75	2.5	4.0	143	179	208
1-14	87	92	74	1.5	1.5	131	125	202
1-17	107	NA	NA	1.7	1.5	133	136	216
1-20	123	NA	NA	1.6	1.5	134	128	179
1-23	139	NA	NA	1.6	2.0	136	128	180
1-26	156	94	71	2.0	1.5	134	124	213
1-29	178	96	69	1.9	2.0	139	126	201
2-2	197	101	74	1.9	1.5	136	120	206
2-7	222	103	71	2.2	2.0	141	154	193
2-10	242	105	69	1.9	2.0	137	154	194
2-13	259	105	68	1.8	2.5	140	136	214
2-16	277	104	68	2.1	2.0	142	144	221
2-10	295	NA	NA	2.1	1.5	134	141	170
2-22	314	105	68	2.5	2.0	136	145	173
2-25	331	105	68	1.7	1.5	127	141	175
2-28	350	105	67	2.0	1.5	129	141	172
3-3	365	105	69	1.8	3.0	128	152	177

Table A-22. VOG Product Water THM's and TOXFP Data Summary[1]

Sample Date	THM's[2]	TOXFP[3]
1-5	40	92
1-8	47	81
1-11	36	120
1-14	13	19
1-17	44	23
1-20	22	38
1-23	20	15
1-26	16	41
1-29	18	49
2-2	15	71
2-7	16	38
2-10	13	39
2-13	30	34
2-16	27	NA
2-19	22	35
2-22	38	36
2-25	32	36
2-28	30	30
3-3	32	40

[1] Incubation conditions: 7 mg/L Cl_2, 96 hrs, 25°C, pH = 8.0

[2] All values are μg/L as individual species.

[3] All values are μg/L as Cl.

Table A-23. VOG Percent Parameter Removal in Product Water (as referenced to the Feed Water)

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (%)	Color (%)	Alkalinity (%)	Total Hardness (%)	Total Dissolved Solids (%)
1-5	23	91	75	95	99	57	59	60
1-8	42	91	75	95	98	56	63	67
1-11	64	91	75	92	97	55	52	63
1-14	84	92	74	95	99	57	63	63
1-17	104	NA	NA	NA	NA	NA	NA	NA
1-20	120	NA	NA	NA	NA	NA	NA	NA
1-23	136	NA	NA	NA	NA	NA	NA	NA
1-26	153	94	71	93	99	57	64	59
1-29	175	96	69	94	98	55	64	62
2-2	194	101	74	94	98	53	63	59
2-7	219	103	71	93	98	51	55	61
2-10	239	105	69	94	98	55	56	60
2-13	256	105	68	NA	NA	NA	NA	NA
2-16	274	104	68	92	98	54	54	57
2-19	292	NA	NA	NA	NA	NA	NA	NA
2-22	311	105	68	92	98	52	56	61
2-25	328	105	68	94	98	51	58	61
2-28	347	105	67	94	98	54	56	63
3-3	362	105	69	94	97	47	54	61

TABLE A-24. VOG Percent Parameter Removal in Product Water (as referenced to the Raw Water)

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (%)	Color (%)	Alkalinity (%)	Total Hardness (%)	Total Dissolved Solids (%)
1-5	23	91	75	88	98	40	42	42
1-8	42	91	75	88	97	34	50	40
1-11	64	91	75	84	93	37	37	50
1-14	84	92	74	90	97	40	45	51
1-17	104	NA	NA	89	97	40	44	40
1-20	120	NA	NA	91	98	41	44	43
1-23	136	NA	NA	91	97	40	43	48
1-26	153	94	71	87	97	41	49	44
1-29	175	96	69	86	96	37	47	45
2-2	194	101	74	89	97	40	48	42
2-7	419	103	71	87	97	36	42	47
2-10	239	105	69	89	97	39	41	49
2-13	256	105	68	89	96	41	44	45
2-16	274	104	68	86	97	37	43	44
2-19	292	NA	NA	NA	NA	NA	NA	NA
2-22	311	105	68	85	97	38	39	46
2-25	328	105	68	90	97	37	44	51
2-23	347	105	67	88	97	36	41	53
3-3	362	105	69	89	95	37	40	48

Table A-25. AID Test Matrix - Product and Raw Water Quality

Feed Pressure (psi)	Product Recovery Rate (%)	Dissolved Organic Carbon (mg/L as C)	Total Dissolved Solids (mg/L)	Total Hardness (mg/L as $CaCO_3$)	True Color (cpu)	Trihalomethane Formation Potential (µg/L)
80	50	1.9	247	153	1.5	42
80	60	2.3	268	174	1.5	36
80	75	2.3	292	189	2.5	32
80	90	2.4	315	194	1.5	29
90	50	1.8	222	154	2.0	23
90	60	2.2	261	173	1.0	25
90	75	2.3	317	187	1.0	30
90	90	2.5	328	194	1.0	30
100	50	2.1	249	152	1.0	18
100	60	2.4	260	169	1.5	29
100	75	2.3	301	187	1.5	33
100	90	2.5	300	187	1.5	49
110	50	2.2	221	144	1.0	29
110	60	2.2	256	168	1.0	40
110	75	2.7	282	187	1.0	43
110	90	2.4	289	202	1.0	60
Raw #1	--	16.9	484	317	36	NA
Raw #2	--	15.4	489	320	37	NA

Table A-26. AID Test Matrix - Percent Parameter Removal Based on Raw Water.

Feed Pressure (psi)	Product Recovery Rate (%)	Dissolved Organic Carbon (%)	Total Dissolved Solids (%)	Total Hardness (%)	True Color (%)
80	50	88	49	52	96
80	60	86	45	45	96
80	75	86	40	41	93
80	90	85	35	39	96
90	50	89	54	52	95
90	60	86	46	46	97
90	75	86	35	41	97
90	90	85	33	39	97
100	50	87	49	52	97
100	60	85	47	47	96
100	75	86	38	41	96
100	90	85	38	41	96
110	50	86	55	55	97
110	60	86	47	47	97
110	75	83	42	41	97
110	90	85	41	37	97

Table A-27. ACME Improvement District Pilot Plant Operating Data Summary

Date	Time of Operation (hr:min)	Elapsed Time of Operation (hr:min)	Product Flux (gals./day-ft²)	Feed Pressure (psi)	Feed Water Temperature (°C)
3-12	4:00	4:00	NA	NA	NA
3-13	11:45	15:45	NA	NA	NA
3-14	5:40	21:25	NA	NA	NA
3-15	7:20	28:45	13.6	90	25
3-16	14:40	43:25	13.8	91	25
3-17	14:45	58:05	13.6	91	25
3-18	6:25	64:30	13.6	91	25
3-19	10:45	75:15	13.9	92	25
3-20	14:30	89:45	13.9	91	25
3-21	14:00	103:45	13.9	92	25
3-22	14:05	117:50	13.9	90	25
3-23	11:35	129:25	13.9	92	25
3-24	14:30	143:55	13.9	91	25
3-25	14:15	158:10	14.1	90	25
3-26	13:45	171:55	14.2	91	24.6
3-27	14:35	186:30	14.2	91	24.3
3-28	14:15	200:45	14.2	91	24.4
3-29	14:15	215:00	14.2	91	24.5
3-30	13:30	228:30	14.2	91	24.5
3-31	14:30	243:00	14.2	90	24.8
4-1	14:30	257:30	14.2	90	24.9
4-2	14:35	272:05	14.2	91	24.0
4-3	14:15	286:20	14.4	91	24.5
4-4	11:15	297:35	14.2	90	24.5
4-5	14:15	311:50	14.4	90	24.7
4-6	14:35	326:25	14.4	90	24.6
4-7	14:55	341:20	14.2	90	24.7
4-8	14:20	356:40	14.4	90	24.8
4-9	14:20	371:00	14.9	100	24.4
4-10	1:15	372:15	NA	NA	NA
4-11	14:45	387:00	14.9	99	24.3
4-12	9:45	396:45	15.0	100	24.5
4-13	14:45	411:30	15.2	100	24.2
4-14	15:10	426:40	15.4	100	24.7
4-15	14:20	441:00	15.2	100	23.7
4-16	14:10	455:10	15.2	100	24.4
4-17	14:35	469:45	15.4	100	24.6
4-18	14:40	484:25	15.4	98	24.5
4-19	14:45	499:10	15.2	100	24.8
4-20	14:50	514:00	15.2	100	24.8
4-21	14:55	528:55	15.2	100	24.8
4-22	13:50	542:45	15.7	100	24.8
4-23	14:10	556:55	15.2	100	24.8

(continued)

Table A-27 (continued). ACME Improvement District Pilot Plant Operating Data Summary

Date	Time of Operation (hr:min)	Elapsed Time of Operation (hr:min)	Product Flux (gals./day-ft^2)	Feed Pressure (psi)	Feed Water Temperature (°C)
4-24	14:30	571:25	15.4	100	25.0
4-25	14:15	585:40	15.7	100	25.0
4-26	14:35	600:15	15.4	99	25.0
4-27	15:05	615:20	15.2	99	25.0
4-28	15:15	630:35	15.4	99	25.0
4-29	10:00	640:35	15.5	99	25.0
4-30	9:40	650:15	15.4	100	24.9
5-1	8:25	658:40	15.4	99	24.9
5-2	13:00	671:40	15.5	100	25.0
5-3	14:25	686:05	15.5	100	25.1
5-4	14:15	700:20	15.4	100	25.0
5-5	14:15	714:35	15.4	100	24.5
5-6	10:00	724:35	20.8	100	25.0
5-7	14:20	738:55	20.5	100	24.8
5-8	14:35	753:30	20.5	100	24.9
5-9	14:20	767:50	20.5	100	24.9
5-10	13:35	781:25	20.5	100	24.6
5-11	8:35	790:00	20.5	100	25.2
5-12	14:40	804:40	20.5	100	25.0
5-13	12:20	817:00	20.5	100	25.1
5-14	8:00	825:00	20.5	99	25.1
5-15	14:35	839:35	20.5	100	25.2
5-16	11:30	851:05	20.5	101	25.0
5-17	10:00	861:05	20.5	101	25.0
5-18	14:15	875:20	20.2	101	25.0
5-19	13:35	888:55	20.3	101	25.0
5-20	00:45	889:40	20.6	102	25.0
5-21	6:45	896:25	20.2	103	25.0
5-22	7:00	903:25	20.2	100	25.0
5-23	7:15	910:40	20.0	100	25.0
5-24	6:15	916:55	20.0	100	25.0
5-25	12:45	929:40	20.0	100	25.0
5-26	11:25	941:05	20.3	100	25.0
5-27	11:10	952:15	20.3	100	25.0
5-28	22:00	974:00	20.0	100	25.0
5-29	14:40	988:40	20.0	99	25.0
5-30	14:50	1003:30	20.0	99	25.0
5-31	14:15	1017:45	20.0	99	

Table A-28. AID Raw Water Quality Data Summary

Sample Date	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
3-17	16.3	44	298	335	528
3-20	15.8	54	307	335	528
3-23	16.1	45	280	312	411
3-26	15.7	55	299	325	461
3-29	14.5	38	292	313	429
4-1	16.5	53	289	313	501
4-4	13.7	38	286	313	444
4-7	14.4	35	281	313	501
4-11	15.9	44	278	313	438
4-14	16.4	51	284	325	439
4-17	15.0	43	283	315	548
4-20	15.5	58	290	315	561
4-23	14.3	46	285	304	411
4-26	14.6	47	290	308	490
4-30	14.2	44	285	329	547
5-3	14.9	52	277	323	554
5-6	15.0	35	277	310	408
5-9	13.6	44	298	318	634
5-12	14.5	56	290	326	596
5-15	15.8	38	288	310	468
5-18	14.1	37	284	314	501
5-21	13.2	39	284	327	480
5-24	13.0	38	276	325	459
5-27	13.1	36	276	315	464
5-31	12.5	39	268	317	447

Table A-29. AID Feed Water Quality Data Summary

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
3-17	49	91	69	27.6	68	432	446	725
3-20	81	91	69	28.9	70	374	427	700
3-23	120	92	69	29.0	70	348	402	557
3-26	149	91	68	24.1	80	368	413	608
3-29	207	91	68	22.4	55	351	407	605
4-1	249	90	68	30.0	75	363	421	631
4-4	292	90	68	32.2	58	353	408	587
4-7	332	90	68	25.2	64	357	412	632
4-11	377	99	79	30.7	70	356	412	574
4-14	416	100	83	28.1	75	367	434	604
4-17	461	100	83	27.5	75	359	412	590
4-20	505	100	83	28.8	75	363	415	--
4-23	549	100	83	26.8	65	361	398	647
4-26	591	99	83	27.0	75	361	402	620
4-30	646	100	83	28.8	60	365	423	662
5-3	676	100	83	29.8	70	353	423	684
5-6	720	100	76	27.1	45	344	412	599
5-9	759	100	76	28.7	80	399	470	849
5-12	796	100	76	32.4	85	399	462	789
5-15	830	100	76	34.7	70	418	449	595
5-18	866	101	75	34.6	70	399	438	628
5-21	895	103	75	33.6	75	382	449	659
5-24	916	100	75	32.3	70	380	453	664
5-27	947	100	76	--	--	--	--	--
5-31	1009	99	75	32.9	70	376	455	676

Table A-30. AID Reject Water Quality Data Summary

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
3-17	49	91	69	52.2	110	518	659	1005
3-20	81	91	69	54.6	125	636	644	1053
3-23	120	92	69	51.5	120	520	593	857
3-26	149	91	68	43.2	135	529	623	924
3-29	207	91	68	49.4	110	521	612	909
4-1	249	90	68	52.7	120	529	655	1002
4-4	292	90	68	52.1	105	542	614	1012
4-7	332	90	68	45.7	110	529	621	1009
4-11	377	99	79	50.3	120	529	640	936
4-14	416	100	83	50.5	130	575	668	1006
4-17	461	100	83	51.9	125	550	648	1103
4-20	505	100	83	53.5	140	556	659	1003
4-23	549	100	83	57.6	120	575	651	1044
4-26	591	99	83	55.1	130	566	661	1081
4-30	646	100	83	57.0	125	567	680	1133
5-3	676	100	83	56.2	135	546	682	1158
5-6	720	100	76	50.0	110	546	876	901
5-9	759	100	76	53.3	130	588	712	1193
5-12	796	100	76	55.5	150	575	736	1100
5-15	830	100	76	65.0	150	583	688	1031
5-18	866	101	75	62.6	145	587	688	1102
5-21	895	103	75	61.8	150	552	699	1022
5-24	916	100	75	59.6	150	560	691	1011
5-27	947	100	76	--	--	--	--	--
5-30	1009	99	75	61.2	150	560	672	1036

Table A-31. AID Product Water Quality Data Summary

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (mg/L as C)	Color (cpu)	Alkalinity (mg/L as $CaCO_3$)	Total Hardness (mg/L as $CaCO_3$)	Total Dissolved Solids (mg/L)
3-17	49	91	69	2.3	1.5	181	194	323
3-20	81	91	69	2.6	2.5	186	206	318
3-23	120	92	69	2.7	1.5	176	181	218
3-26	149	91	68	2.4	1.5	185	181	240
3-29	207	91	68	2.4	1.0	172	177	237
4-1	249	90	68	2.4	1.0	179	177	304
4-4	292	90	68	2.4	1.0	180	177	254
4-7	332	90	68	2.5	1.0	172	181	241
4-11	377	99	79	2.0	1.5	173	179	240
4-14	416	100	83	1.6	1.5	181	181	245
4-17	461	100	83	2.5	0.5	170	177	286
4-20	505	100	83	2.4	0.5	170	172	322
4-23	549	100	83	2.0	1.5	180	173	288
4-26	591	99	83	2.2	1.5	185	175	318
4-30	646	100	83	2.8	1.0	168	189	305
5-3	676	100	83	2.7	1.0	176	174	314
5-6	720	100	76	2.9	1.0	170	177	278
5-9	759	100	76	2.6	1.5	197	207	338
5-12	796	100	76	2.2	0.5	193	199	276
5-15	830	100	76	1.2	1.5	188	198	288
5-18	866	101	75	1.5	1.5	192	202	323
5-21	895	103	75	1.2	1.5	180	202	261
5-24	916	100	75	2.3	2.0	184	204	256
5-27	947	100	76	1.6	2.0	184	202	268
5-31	1009	99	75	1.6	2.5	180	204	283

Table A-32. AID Product Water THM's and TOXFP Data Summary[a]

Sample Date	Chlorine Dosage			
	7 ppm as Cl_2		10 ppm as Cl_2	
	THM's[b]	TOXFP[c]	THM's[b]	TOXFP[c]
3-17	NA	NA	50	68
3-20	NA	NA	72	32
3-23	24	27	NA	NA
3-26	19	19	NA	NA
3-29	18[d]	NA	NA	NA
4-1	18[d]	22[d]	NA	NA
4-4	NA	NA	40	29
4-7	NA	NA	23	26
4-11	22	20	23	30
4-14	28	11	27	27
4-17	NA	NA	36	30
4-2-	NA	NA	38	28
4-23	NA	NA	25	34
4-26	NA	NA	42	32
4-3-	NA	NA	64	34
5-3	NA	NA	34	42
5-6	NA	NA	33	34
5-9	77	78	113	58
5-12	69	55	74	52
5-15	47	40	57	56
5-18	50	34	54	51
5-21	48	46	45	52
5-24	39	52	73	64
5-27	NA	NA	74	55
5-31	26	27	32	36

[a] Incubation conditions: 96 hrs, 25°C, pH = 8.0

[b] All THM's values are µg/L as the individual species.

[c] All TOXFP values are µg/L as Cl.

[d] Measured chlorine residual was 0.0 mg/L.

Table A-33. AID Percent Parameter Removal in Product Water (as referenced to the Feed Water)

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (%)	Color (%)	Alkalinity (%)	Total Hardness (%)	Total Dissolved Solids (%)
3-17	49	91	69	92	98	58	57	55
3-20	81	91	69	91	96	50	52	55
3-23	120	92	69	91	98	49	55	61
3-26	149	91	68	90	98	50	56	61
3-29	207	91	68	89	98	51	57	61
4-1	249	90	68	92	99	51	58	52
4-4	292	90	68	93	98	49	57	57
4-7	332	90	68	90	98	52	56	62
4-11	377	99	79	93	98	51	56	58
4-14	416	100	83	94	98	51	58	59
4-17	461	100	83	90	99	53	57	52
4-20	505	100	83	92	99	53	59	46
4-23	549	100	83	93	98	50	57	55
4-26	591	99	83	92	98	49	56	49
4-30	646	100	83	90	98	54	55	54
5-3	676	100	83	91	99	50	59	54
5-6	720	100	76	89	98	51	57	54
5-9	759	100	76	91	98	51	56	60
5-12	796	100	76	93	99	52	57	65
5-15	830	100	76	96	98	55	56	52
5-18	866	101	75	97	98	52	54	49
5-21	895	103	75	95	98	53	55	60
5-24	916	100	75	93	97	52	55	61
5-31	1009	99	75	95		52	55	58

TABLE A-34. AID Percent Parameter Removal in Product Water (as referenced to the Raw Water)

Sample Date	Elapsed Time (hr)	Feed Pressure (psi)	Product Water Recovery (%)	DOC (%)	Color (%)	Alkalinity (%)	Total Hardness (%)	Total Dissolved Solids (%)
3-17	49	91	69	86	97	39	42	39
3-20	81	91	69	84	95	39	39	40
3-23	120	92	69	83	97	37	42	47
3-26	149	91	68	85	97	38	44	48
3-29	207	91	68	83	97	41	43	45
4-1	249	90	68	85	98	38	43	39
4-4	292	90	68	82	97	37	43	43
4-7	332	90	68	83	97	39	42	52
4-11	377	99	79	87	97	38	43	45
4-14	417	100	83	90	97	36	44	44
4-17	461	100	83	83	99	40	44	48
4-20	505	100	83	85	99	41	45	43
4-23	549	100	83	86	97	37	43	30
4-26	591	99	83	85	97	36	46	35
4-30	646	100	83	80	98	41	43	44
5-3	676	100	83	82	98	36	46	43
5-6	720	100	76	81	97	39	43	32
5-9	759	100	76	81	97	34	35	47
5-12	796	100	76	85	94	33	39	54
5-15	830	100	76	92	96	35	36	38
5-18	866	101	75	89	96	32	36	36
5-21	895	103	75	91	96	37	38	46
5-24	916	100	75	82	95	33	37	44
5-31	1009	99	75	88	93	35	35	39

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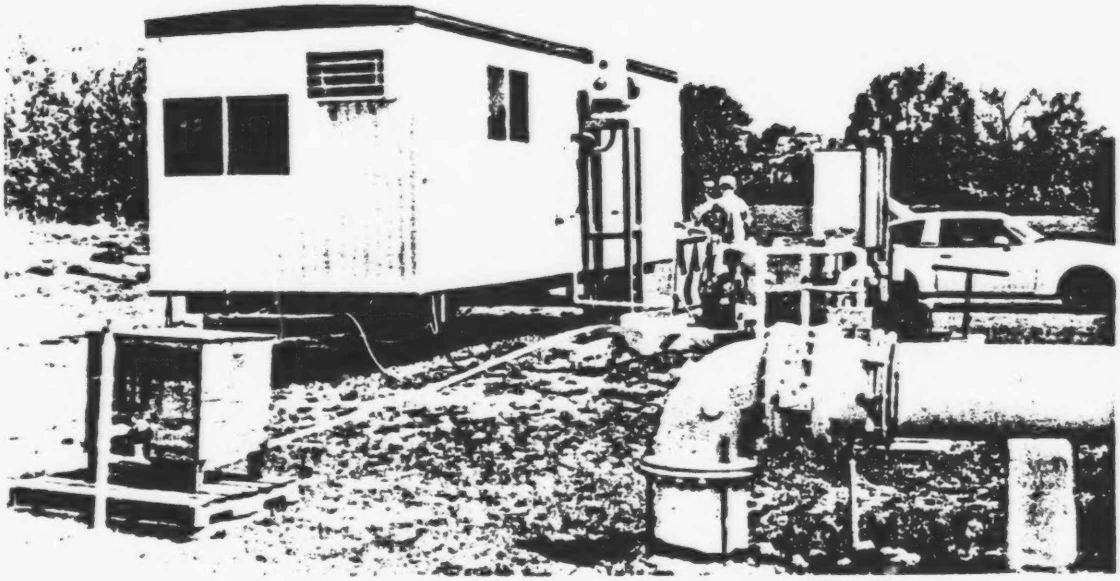

Figure A-1 Membrane Pilot Plant Trailer-Exterior Photograph Showing the Aerator and Booster Pump.

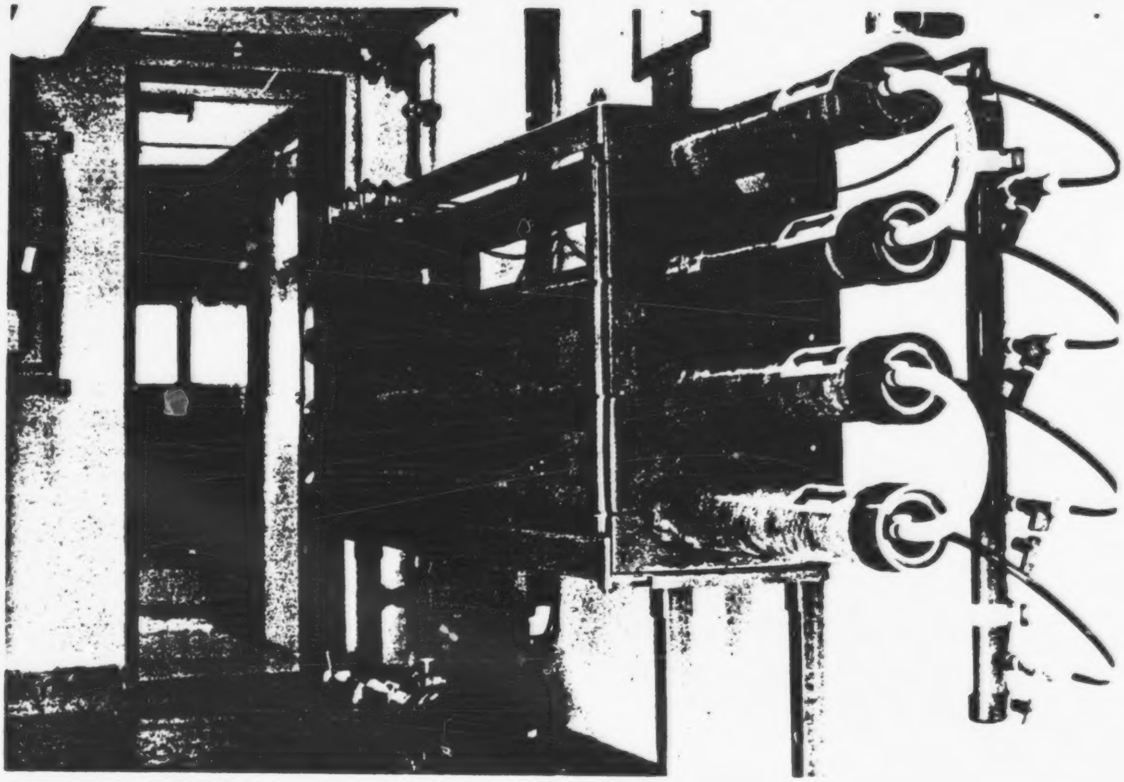

Figure A-2 Membrane Pilot Plant Trailer-Interior Photograph Showing the Pressure Vessels, Product Water Collection Manifold and Tonkaflo Ultrafiltration Pump.

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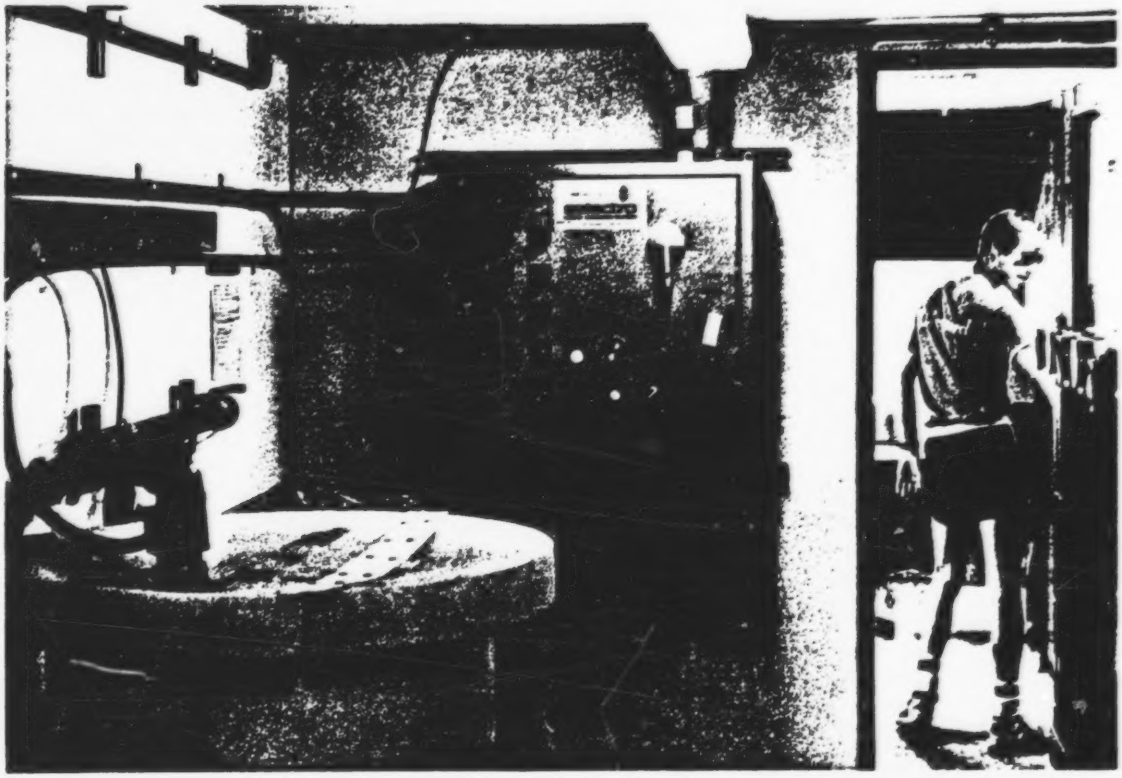

Figure A-3 Membrane Pilot Plant Trailer-Interior Photograph Showing the Control Board.

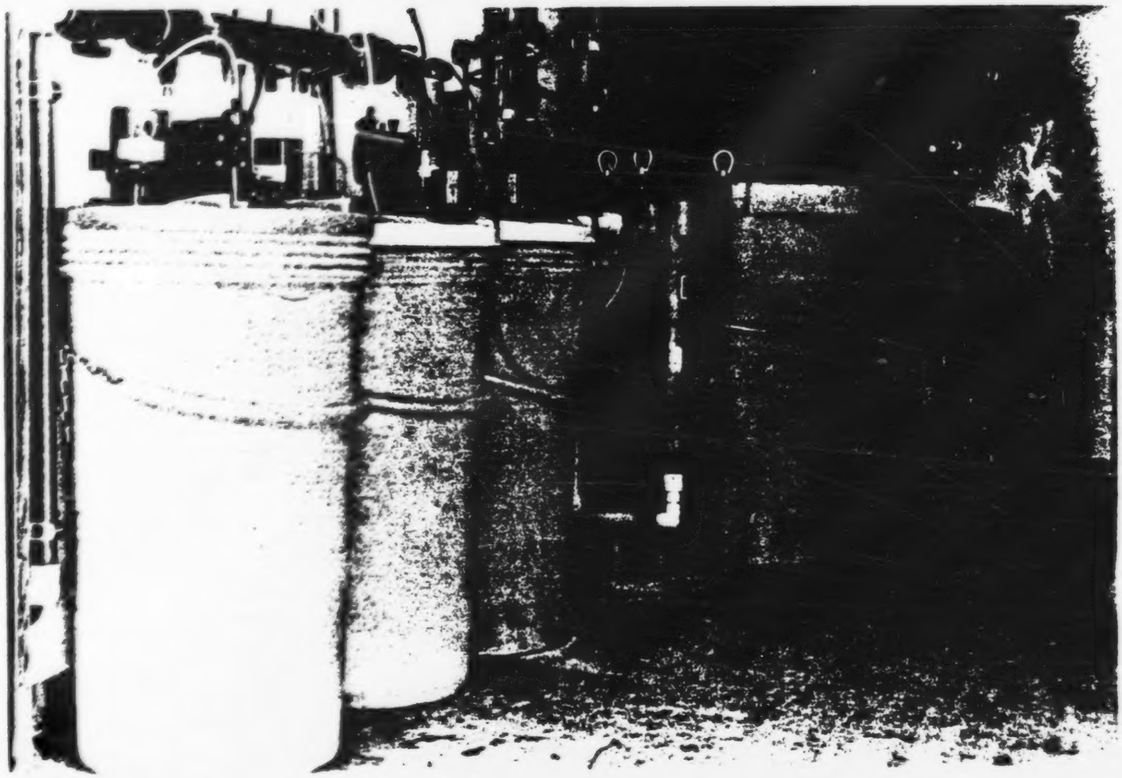

Figure A-4 Membrane Pilot Plant Trailer-Interior Photograph Showing the Chemical Holding Tanks, Pre-filter and Clear Well.

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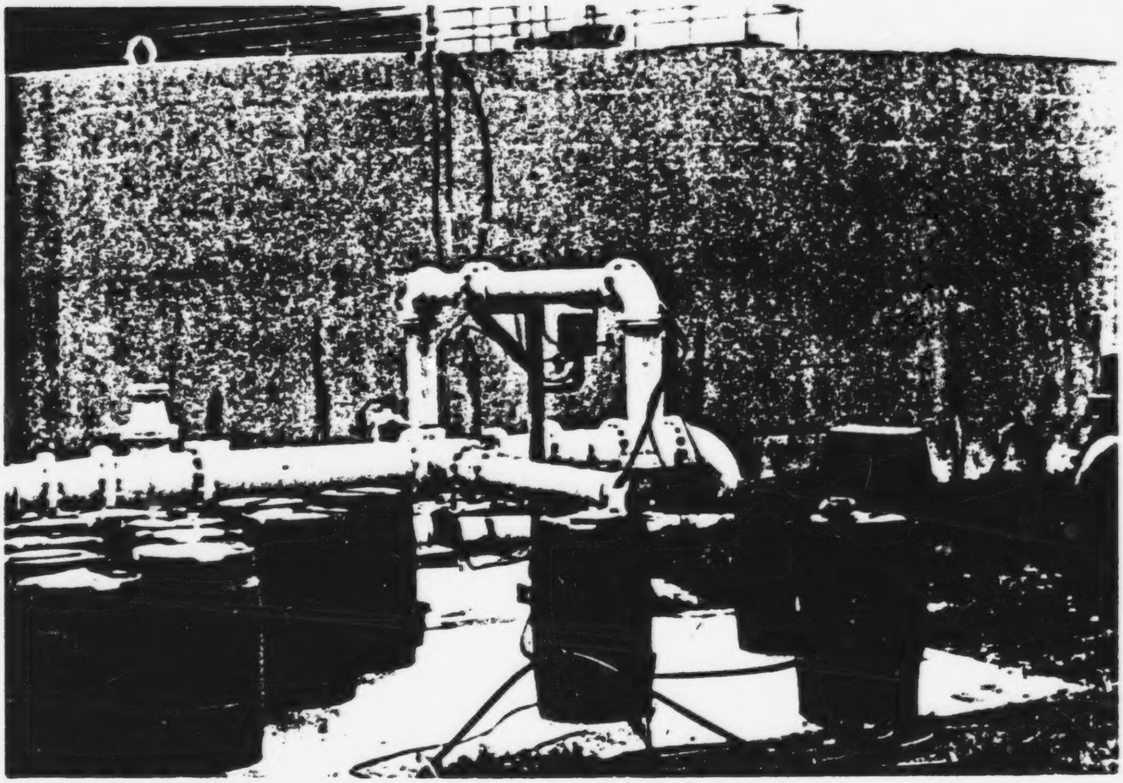

Figure A-5 Acid-Feed Facilities at Olga for PACl/Alum Plant Test.

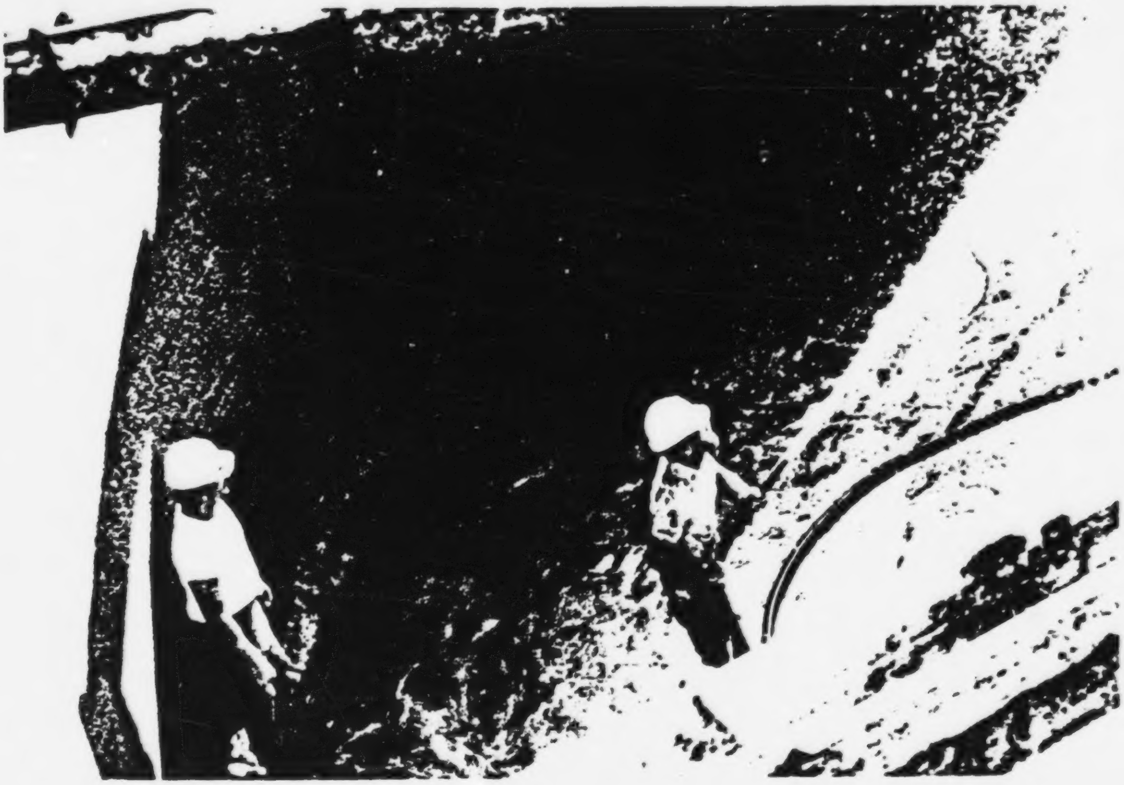

Figure A-6   Cleaning of Reactor for PACl/Alum Plant Test.

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Figure A-7 Soda-Ash Feed Facilities at Olga for PACl/Alum Plant Test.

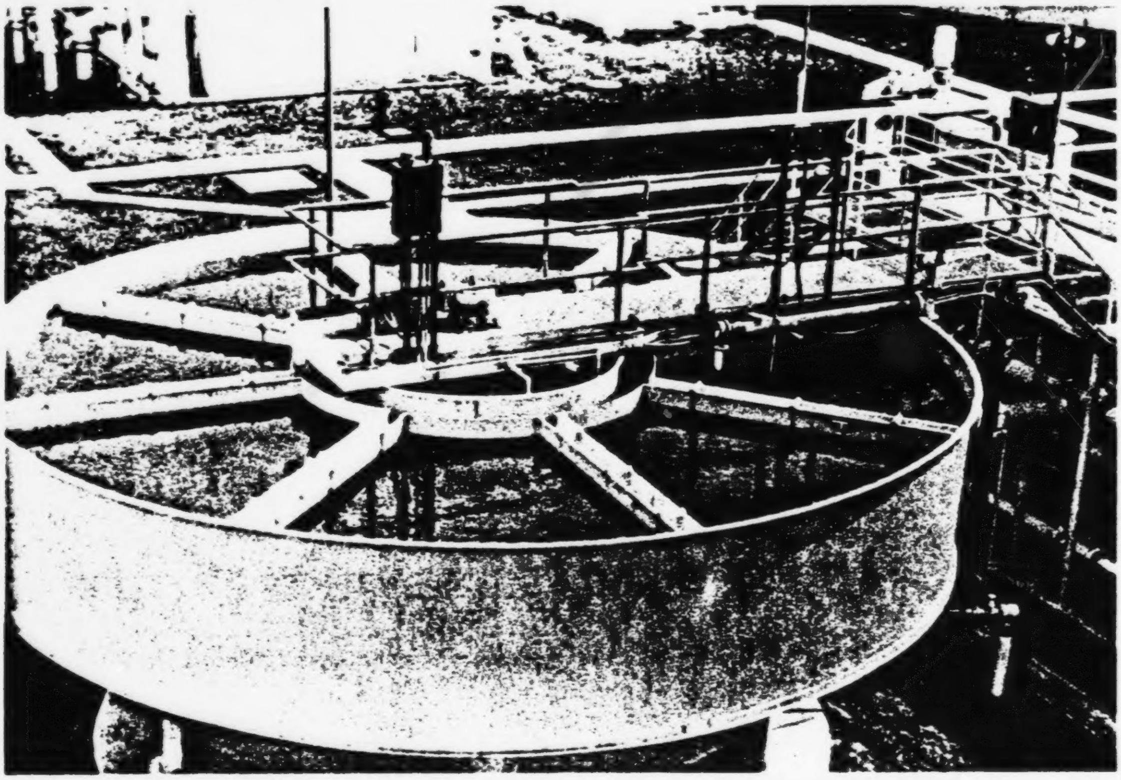

Figure A-8 Lime-Softening Accelator at AID for Sequential Treatment.

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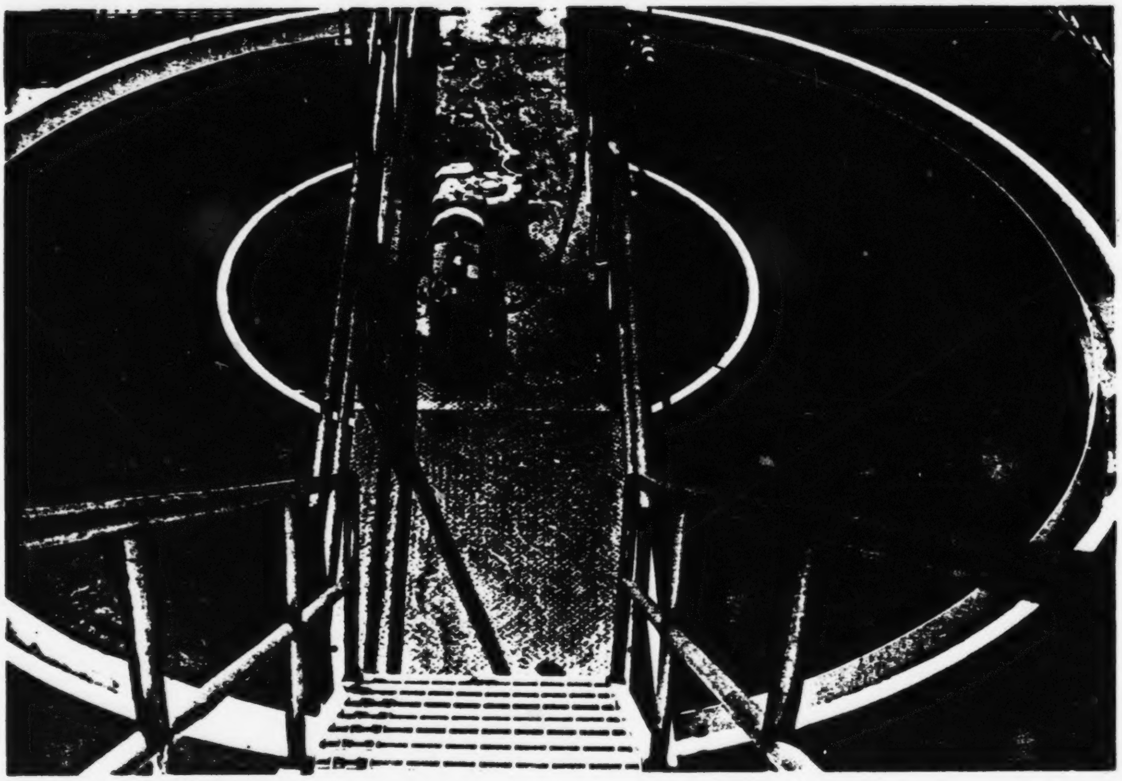

Figure A-9 Alum Coagulation Accelator at AID for Sequential Treatment

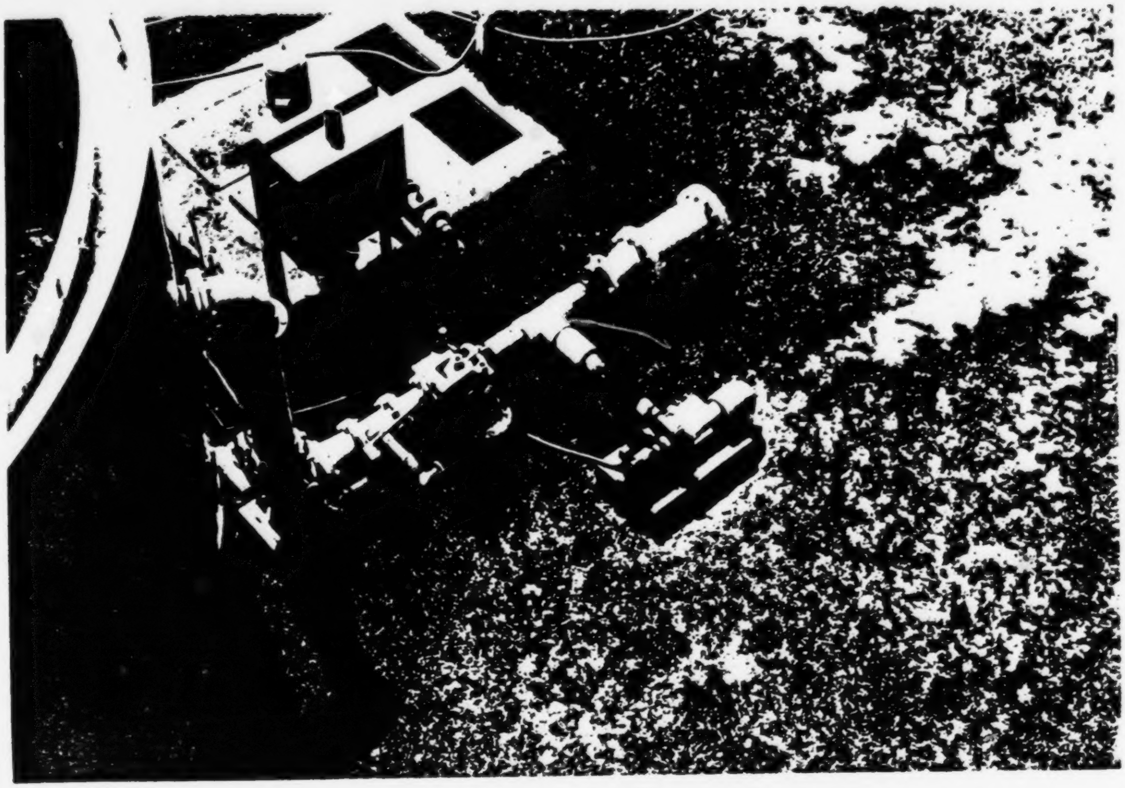

Figure A-10 Pump for Transfer between Lime Softening and Alum Coagulation Reactor.

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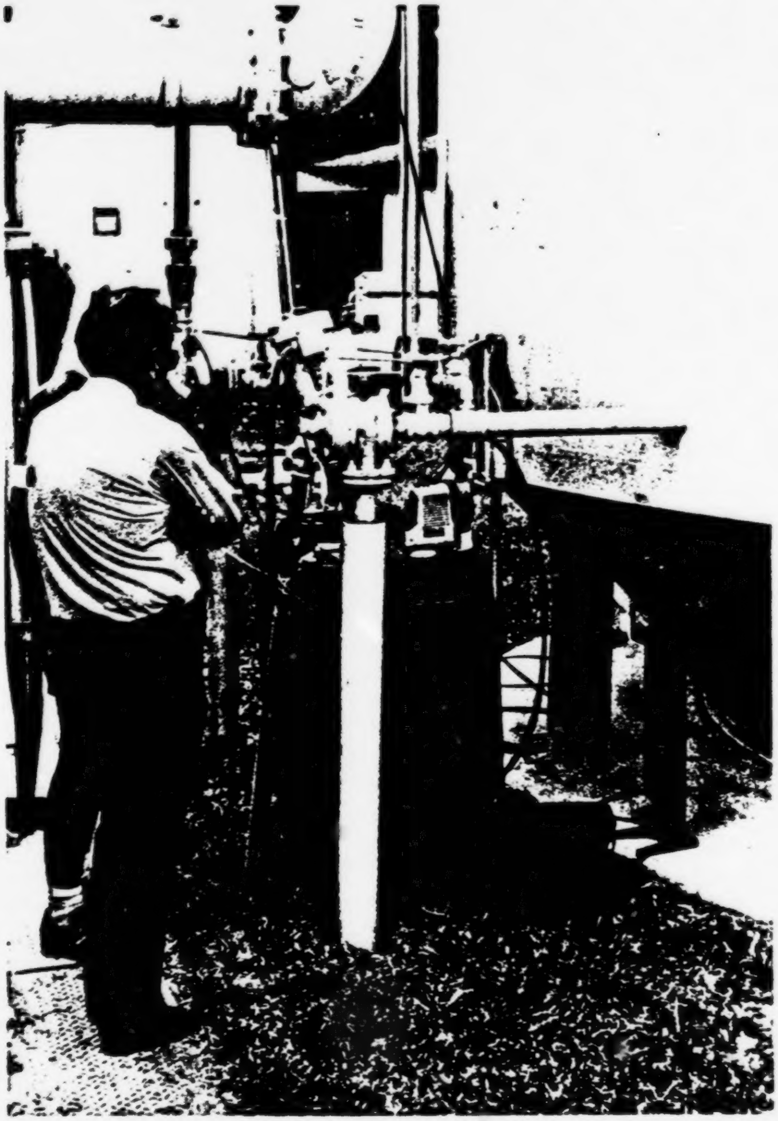

Figure A-11 Chemical Feeders for Sequential Treatment Plant Test.

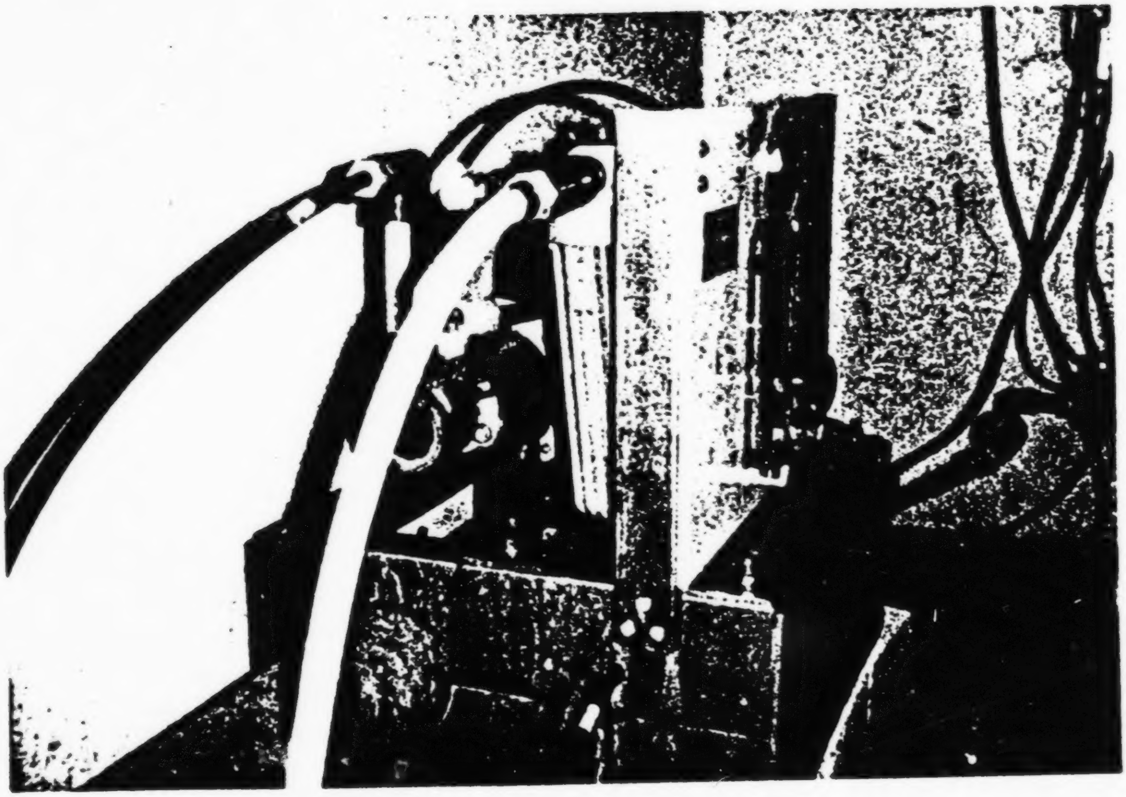

Figure A-12 Modified BTI Selectro "007" Bench-Scale Membrane Unit as Installed at Olga for the Extended Study.

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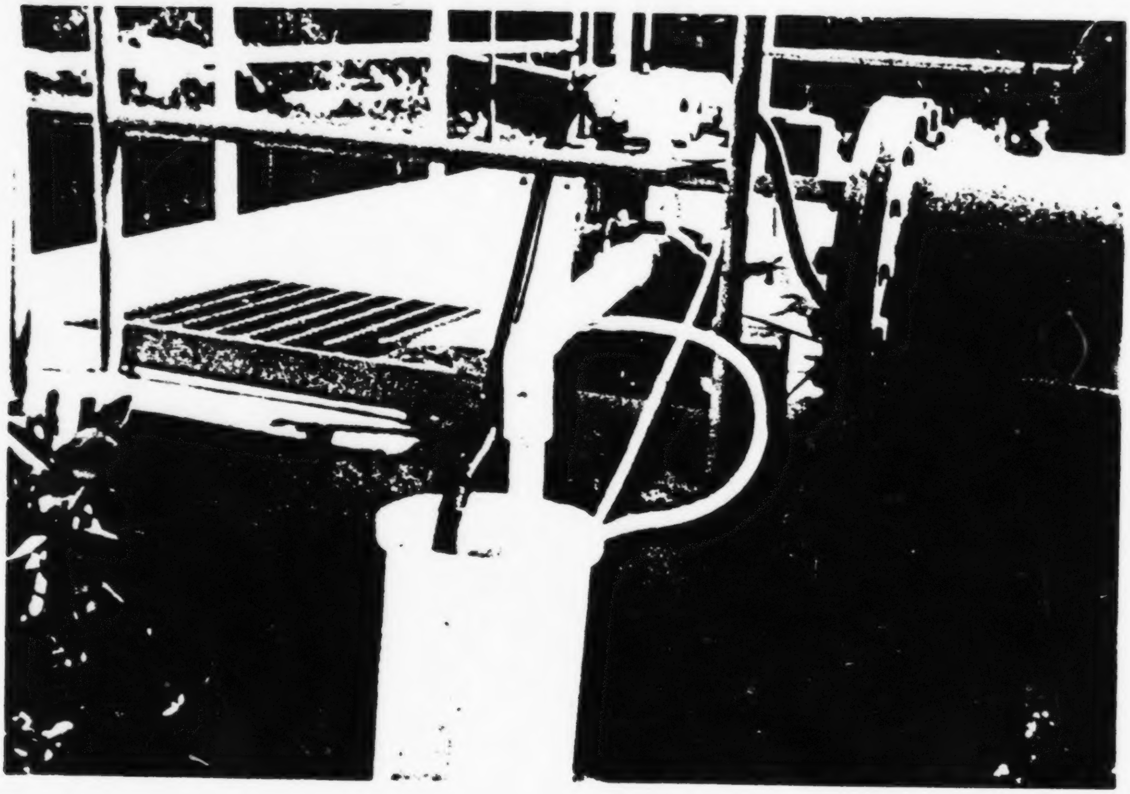

Figure A-13 Raw Water Holding Tank for the Bench-Scale Membrane Unit as Installed at Olga for the Extended Study.

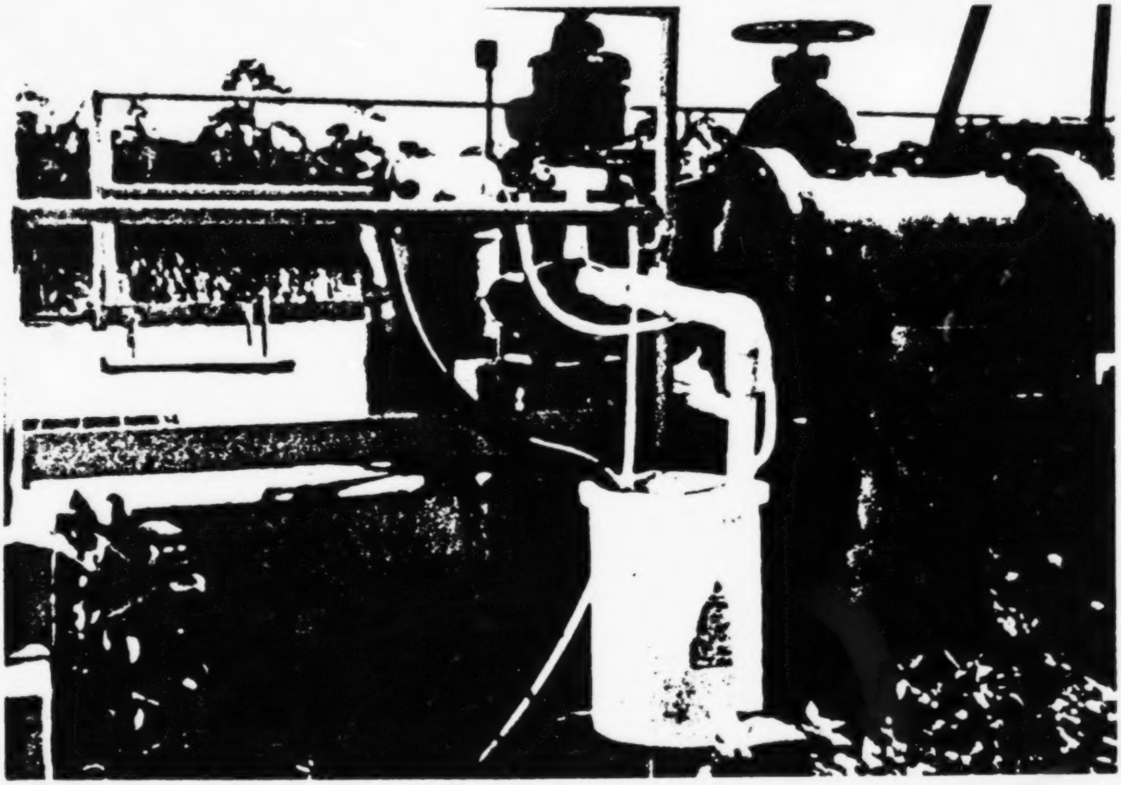

Figure A-14 Sand Filter for the Bench-Scale Membrane Unit at Olga for the Extended Study.

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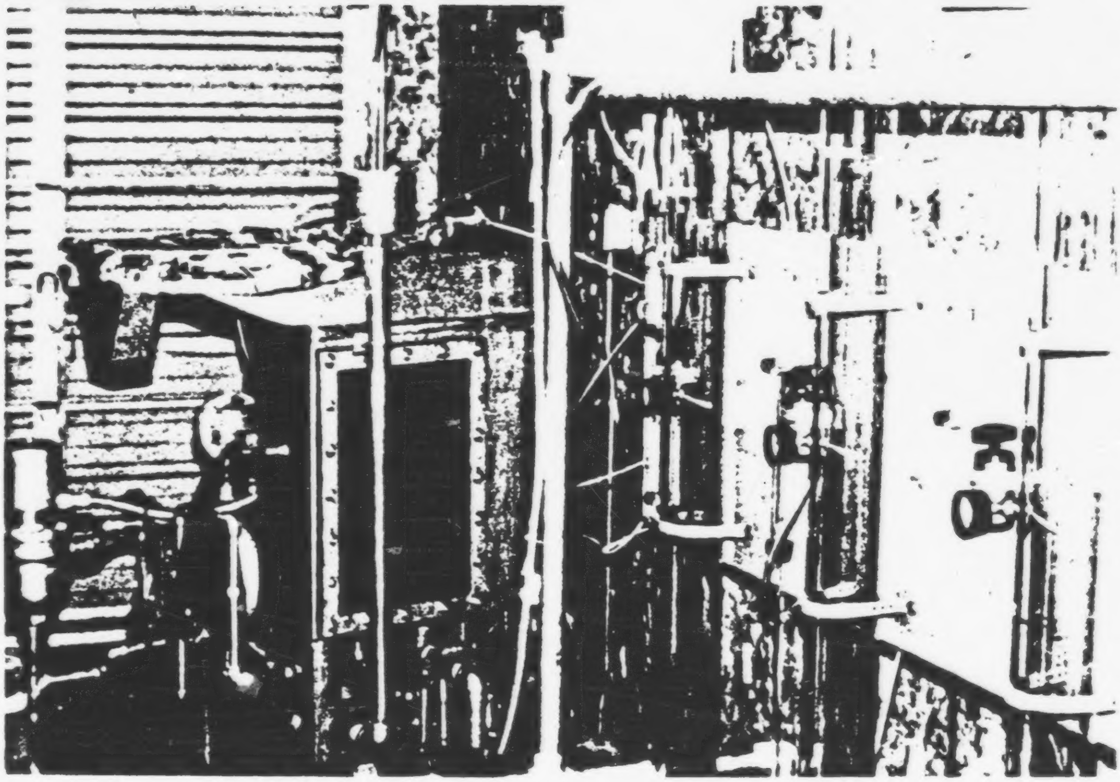

Figure A-15 DAF Pilot Unit in Olga Trailer.

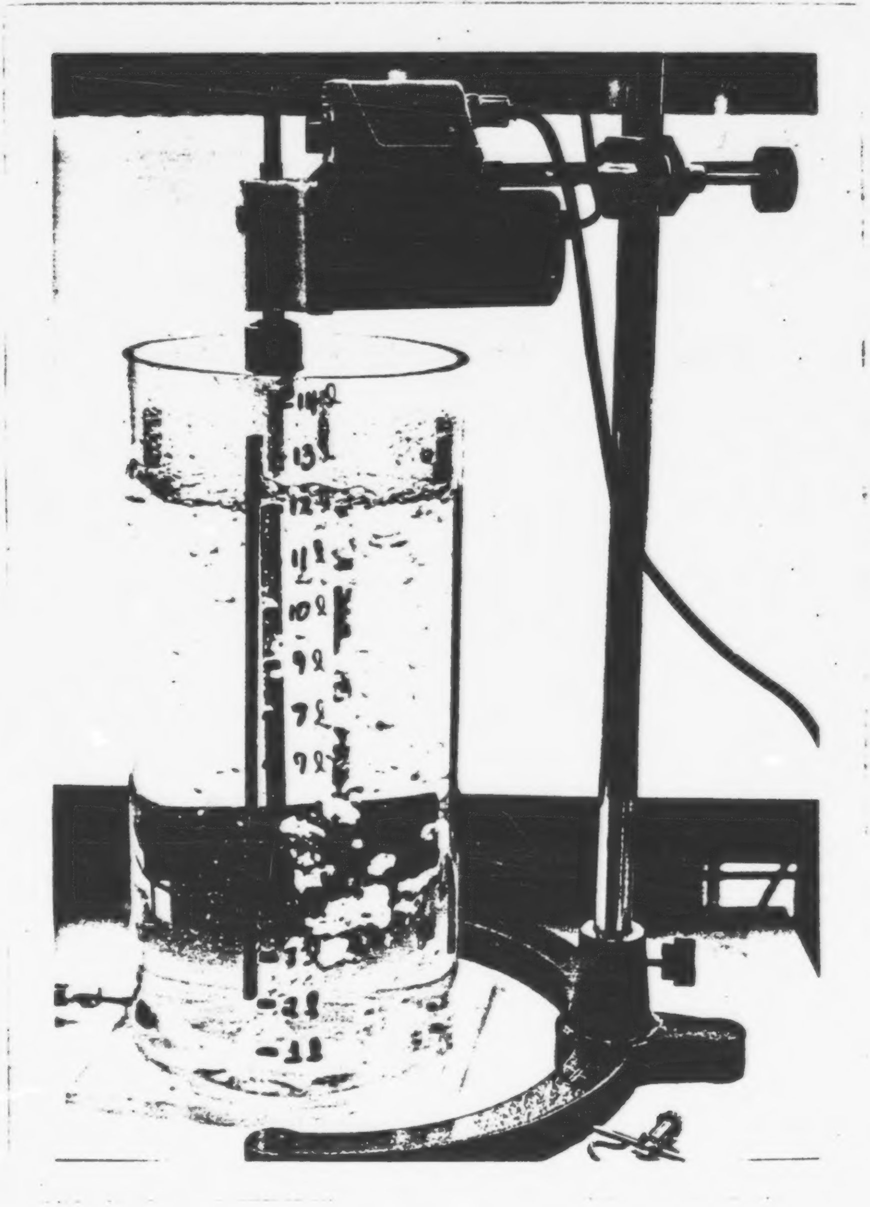

Figure A-16 Mixer and Reactor for a Common Environment Sequential Coagulation and Dissolved Air Flotation.